MARCH 2021 | HydrocarbonProcessing.com

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## PETROCHEMICAL TECHNOLOGY

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**MARCH 2021** | Volume 100 Number 3
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## HP Industry Perspectives

HYDROCARBON PROCESSING IRPC PROCESS

# IRPC Process Technologies: An event designed to showcase advancing technologies in the HPI

The hydrocarbon processing industry (HPI) is steadily advancing through the energy transition. New regulations and company/governmental initiatives have and will drive the HPI towards developing ultra-low-sulfur fuels, alternative fuels, biofuels, green petrochemicals and plastics recycling technologies. These trends are shaping the way companies invest in new research and development programs and where capital-intensive investments are being made in new production capacity.

However, investments in conventional refining and petrochemical processing production routes are still being made globally. Hundreds of billions of dollars are being invested to produce refined products and petrochemicals to satisfy increasing demand. This new capacity is utilizing the latest processing technologies within the HPI.

These two crucial processing routes—conventional and energy transition technologies—will be instrumental in providing the fuels and petrochemical products of the future. Due to the critical nature of these technologies, *Hydrocarbon Processing* will focus on these two processing routes in its next International Refining and Petrochemical Conference (IRPC). IRPC Process Technologies—Process Optimization and Transition Technologies—will examine the challenges affecting the HPI, as well as the opportunities that are emerging in new refining and petrochemical processing technologies. The global, virtual event will be held June 2–3 and will feature dozens of speakers from around the world.

**Agenda and tracks.** The agenda for IRPC Process Technologies is now available at www.HydrocarbonProcessing.com/Events. The event will be kicked-off with Leon de Bruyn, CEO, Lummus Technology, who will discuss how the processing technology licensor is adapting to the current market and how it will adapt/evolve in the future.

The event will feature two primary tracks: One on conventional refining and petrochemical processing technologies and the other on energy transition technologies. The agenda's conventional tracks will focus on distillation, alkylation, desulfurization, ethylene and propylene production, fluid catalytic cracking, hydrocracking, coking and refining-petrochemical integration.

The event's energy transition technologies track will focus on alternative fuels, biofuels, green petrochemicals production, refining and petrochemicals sustainability, circular economy and plastics recycling technologies. HP


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# Petrochemicals: Changing technologies and configurations to satisfy future demand

For nearly a decade, capital investments in new petrochemical capacity have skyrocketed. This significant surge in capital investments (i.e., hundreds of billions of dollars) is in response to sharp increases in demand for plastics and chemicals, especially in developing nations.

However, the global petrochemicals industry has been significantly affected by the COVID-19 pandemic. Several industries were severely hit hard by demand declines (e.g., automotive), leading to a reduction in petrochemical products demand. This predicament has led to many asset owners delaying, deferring and/or abandoning capital-intensive petrochemical plants.

**Global GDP: Is a rebound afoot?** Although the COVID-19 pandemic has decimated fuels and petrochemicals demand and led to sharp declines in nearly every country's gross domestic product (GDP), the International Monetary Fund's January *World Economic Outlook* forecasts global GDP to increase 5.5% in 2021 and 4.2% in 2022. Increases in GDP tend to result in increases in demand for petrochemicals, especially for polyethylene. With forecasted global GDP growth, petrochemical asset owners may see a brighter horizon and revisit plans to restart and/or greenlight additional petrochemical capacity investments.

**Project market share and technological advancements.** At present, *Hydrocarbon Processing's* Construction Boxscore Database is tracking more than 430 active petrochemical products around the world. Nearly half of these projects are in the Asia-Pacific region, followed by the U.S. and the Middle East. These three regions constitute 75% of active petrochemical project market share.

While the U.S. is focusing on processing abundant, cheap shale gas, many nations, especially in Asia and the Middle East, are integrating petrochemical production capacity into existing refining operations and/or investing in capital-intensive grassroots refining and petrochemical integrated mega-complexes. For example, the Construction Boxscore Database is tracking tens of billions of dollars in new refining and petrochemical integrated production capacity in places such as China, India and the Middle East.

However, robust knowledge and detailed engineering is needed to successfully integrate petrochemical production capacity into existing refining assets, as well as in planning, engineering and building a capital-intensive, grassroots integrated complex. The topic of refining and petrochemical integration is on display within this issue's Special Focus section on petrochemicals technology. Refining and petrochemical integration technologies will also be a major focus at *Hydrocarbon Processing's* IRPC Process Technologies global, virtual event.

The increased use and demand for plastics has created an ecological challenge—plastics waste. However, over the past several years, many organizations have invested heavily in creating new technologies to chemically recycle plastics. Should additional investments be available to research, engineer and build new chemical recycling plants, these technologies will aide nations in mitigating plastic waste. It will not completely solve the problem, but it does showcase how the ingenuity of process engineers can see a problem and engineer a solution. This topic will be a major focus in future issues of *Hydrocarbon Processing*.

**Moving forward.** Although the road has been rough over the past year, the resiliency of the processing industry will ensure that global demand for fuels and petrochemicals will be satisfied, and in a much safer and environmentally-friendly way. HP

## INSIDE THIS ISSUE

S. FERNANDES, Yokogawa Corp., Houston, Texas



# A perspective on new technology impacts on the chemical processing industry

Over the past five years, the industrial evolution has been propelled by the same technologies that have dramatically changed our private lives. These include cloud computing, wireless telecommunications, device mobility, algorithmic power, data storage, cybersecurity and robotics.

However, two critical observations can be made regarding the industrial environment. First, people will discover that these information technology (IT) capabilities are augmenting digital solutions that have already been deployed over the past 50 yr. In other words, the industry has been practicing digitization and digitalization for a long time. The recent changes have made these applications more powerful, more agile, more portable, more intelligent and more autonomous. Second, evolution in the chemical industry is gradual; it is incremental due to the intrinsic liabilities that are associated with the operation of industrial plants.

Unlike the dramatic changes in the way we communicate with one another today, industrial applications evolve in a conservative manner that is driven by personal and environmental safety, physical security and cybersecurity, and basic economics that will always be constraints to the optimization process. Furthermore, we can add another key, recent requirement: sustainability.

The chemical industry has witnessed the deployment of many proven technologies at a large scale catalyzed by the technology enablers of Industry 4.0 and smart manufacturing. Not only has cloud computing enabled a reduction in capital expenditure budgets, but it has also facilitated the availability of process models—whether steady-state or dynamic—regardless of the end user's location. An offshore platform operator can refresh their technical skills by running an operator training simulator while at home. A control room technician in Europe can test reactions to the loss of a compressor in a plant in a virtual environment while the plant is being designed and built in Asia—well ahead of its startup.

Control room operators can be trained in central engineering offices by enabling their access to digital twins that reside in the cloud and deploy dynamic, first principle models to represent process units in far-flung locations. Field operation productivity has also increased, while eliminating accidents and errors by employing augmented reality (AR) mobile devices and leveraging wireless instrumentation.

Technology applications can predict the failure of mechanical equipment (e.g., pump cavitation or the potential of off-spec polymer production) by deploying a machine-learning algorithm, which acquires real-time data such as flow, pressure, temperature and vibration from the process.

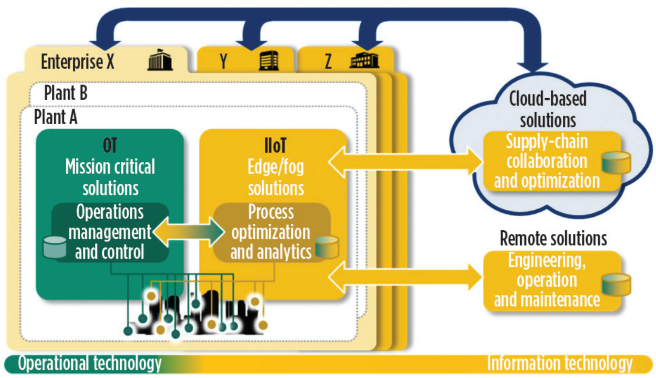


FIG. 1. The convergence of IT with OT: An enterprise architecture.

These platforms are examples of digital twins that are now ubiquitous thanks to the flexibility of the cloud, the fog and the dust. Nevertheless, there are red flags that must be raised regarding these digital applications. For example, a reactor catalyst ages and loses its activity with time; a heat exchanger fouls if not regularly cleaned; a turbine lube oil oxidizes and its viscosity changes if it is not periodically replaced; and a distillation column will weep. In other words, industrial processes are "living" entities. They change with time. Any mathematical representation—such as a digital twin—needs to follow what is happening in the plant, adapt to it, and allow it to be updated by one or more mechanisms; otherwise, it will eventually fall into disuse. Assets require attention; they demand budgets to maintain their sustainability.

This has happened to a field-proven technology that was developed in the early 1970s, which has generated tangible and intangible benefits: advanced process control—particularly multivariable predictive control. However, some plants have this technology turned off after commissioning. This is a warning about a key consideration that cannot be overlooked when implementing new technologies: change. Processes change, people change, technologies change and enterprises change. Sustainable design must be incorporated early in the process.

Nevertheless, the chemical industry continues to look to the future, while embracing the revolution that autonomous operations will bring. The critical need for safer, more reliable, more profitable and more sustainable operations requires a smart balance when deploying human resources in an industrial environment. Risky field operations, repetitive actions, mundane

activities, unnecessary trips to collect data in the field and inspection in hazardous areas can be intelligently addressed by current and future technologies.

**Industrial applications evolve in a conservative manner that is driven by personal and environmental safety, physical security and cybersecurity, and basic economics that will always be constraints to the optimization process.**

The chemical industry has undergone new challenges, some in the last few years and other fairly recently. What is common? Change is constant. Whenever there is a change, there is probably a new optimal point. If change is constant, then adaptation, flexibility, agility and savviness are constants. What about the future? What about disruption?

According to a report from the Recording Industry Association of America (RIAA), vinyl records accounted for $232.1 MM of music sales in 1H 2020 vs. CDs, which brought in only $129.9 MM. Conversely, music streaming increased 12% to $4.8 B during the same period. This is a clear example of digital transformation! We may not see such a dramatic evolution in the chemical industry due to its intrinsic liability nature. Nevertheless, we are optimistic about what is next.

In 1951, Arthur C. Clarke wrote the short story "The Sentinel." That was the inspiration for Stanley Kubrik's "2001: A Space Odyssey," which was released in 1968. We are now in the year 2021 and control room operators do not interact with a computer the same way that Dr. Dave Bowman interacts with the HAL 9000 computer.

However, picture a humanoid sitting next to a person in a control room. In the humanoid's "brain" (i.e., the CPU) resides all mathematical models of the plant. The humanoid has a digital twin of the whole process inside its body in addition to copies of all process flow diagrams, piping and instrumentation diagrams, drawings and manuals for all equipment, as well as other manuals and process diagrams for the plant's operations. The only struggle for the humanoid is to cope with the personality of the operator on that particular shift. Is it reading the operator's lips like HAL 9000 in "2001: A Space Odyssey?" Well, at this time, like The Go-Go's sang, "Our lips are sealed." HP

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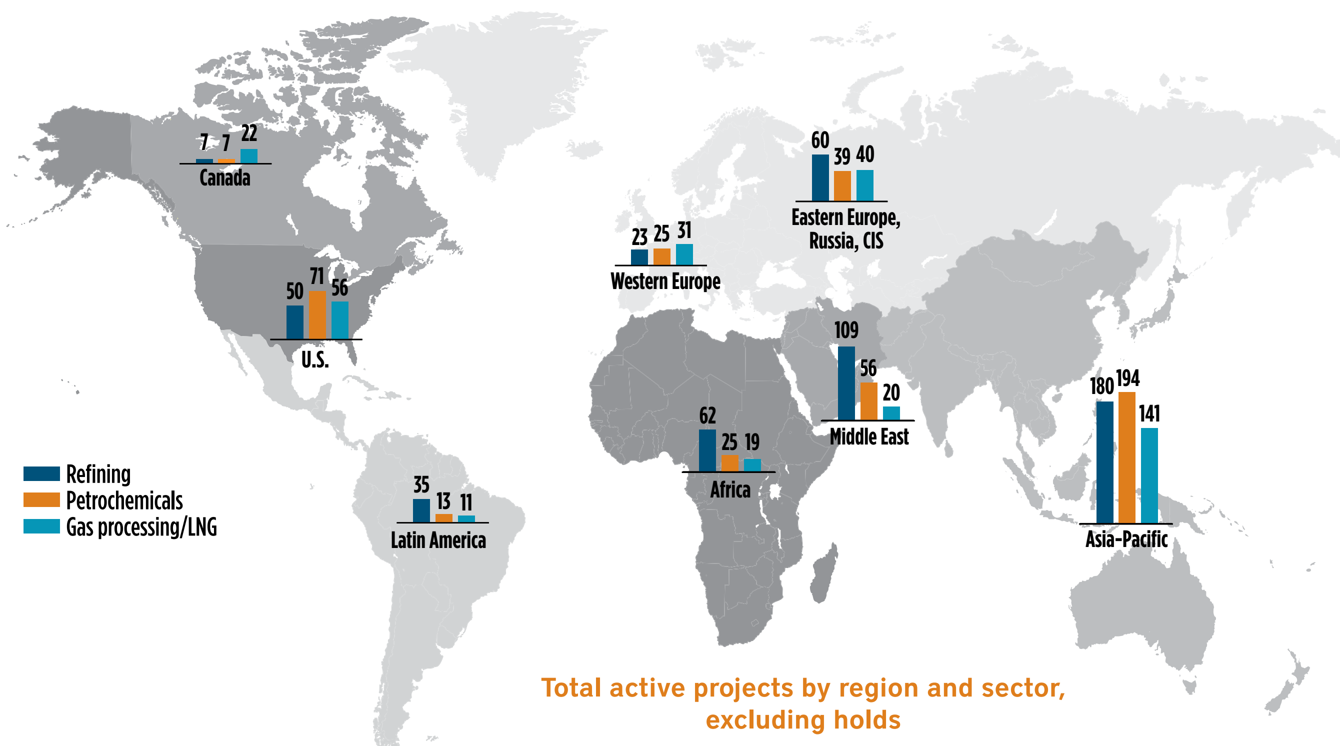
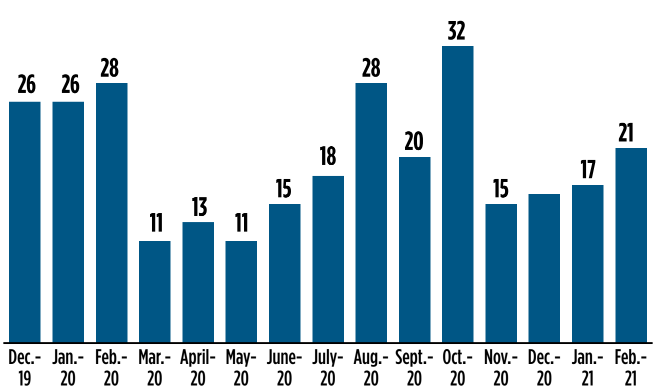
# HP Global Project Data

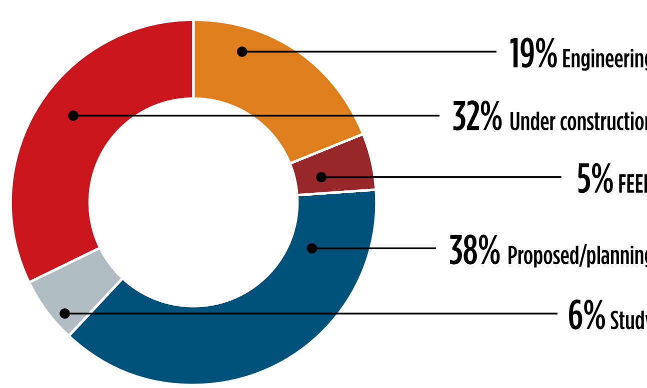
After a tumultuous year, many active capital projects have been deferred, delayed or even abandoned. After excluding projects that have been put on hold, *Hydrocarbon Processing's* Construction Boxscore Database is tracking nearly 1,300 projects around the world in the refining, petrochemicals and gas processing/LNG sectors. At approximately 40%, the Asia-Pacific region is the leader in active projects, followed by the Middle East and the U.S. Asia has also been the leader in new project announcements over the past year. Globally, new project announcements have increased for the past four months. HP




**Total active projects by region and sector, excluding holds**




**New project announcements, December 2019–January 2021**




**Active project market share by activity level**

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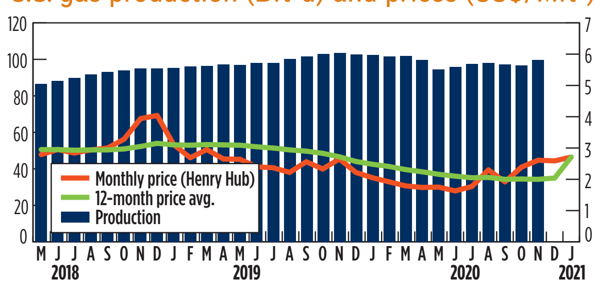
# HP Industry Metrics

U.S. refinery margins extended their upward trend for the fifth consecutive month, exhibiting the largest upturn relative to other regions. European refining economics reversed trend, but saw limited gains relative to other regions given subdued product drawdowns, seasonal weakness and strict COVID-19 related lockdown measures. In Asia, robust performance in light-end markets filtered through to gasoline markets, offsetting poor performance across the mid- and bottom section of the barrel. HP

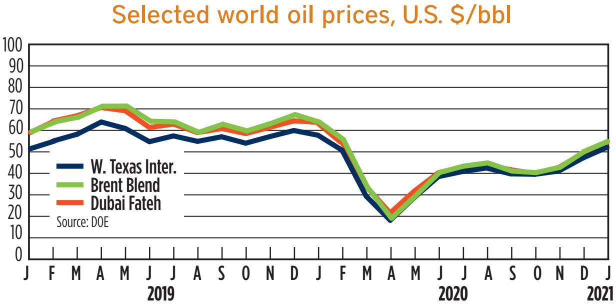
An expanded version of Industry Metrics can be found online at HydrocarbonProcessing.com.

## U.S. gas production ($Bft^3d$) and prices ($US\$/Mft^3$)

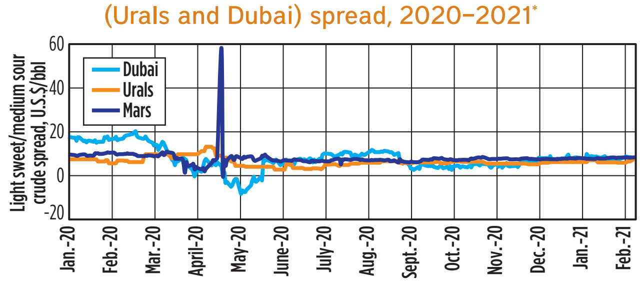



Production equals U.S. marketed production, wet gas. Source: EIA.

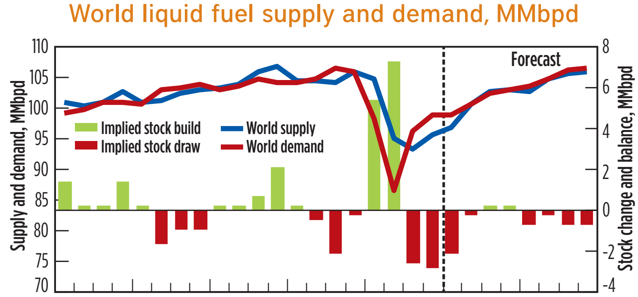
## Selected world oil prices, U.S. $/bbl




## Brent dated vs. sour grades (Urals and Dubai) spread, 2020–2021*

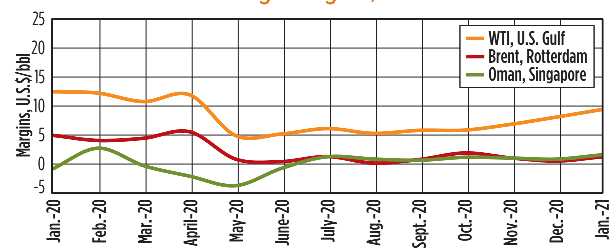



## World liquid fuel supply and demand, MMbpd

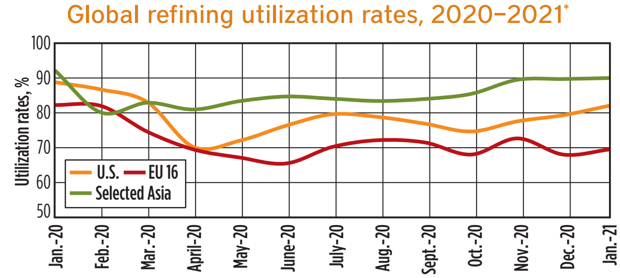



Source: *EIA Short-Term Energy Outlook*, February 2021

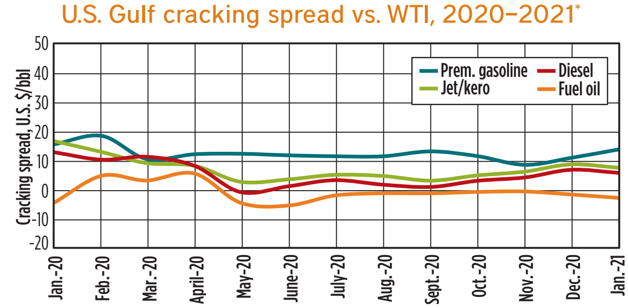
## Global refining margins, 2020–2021*



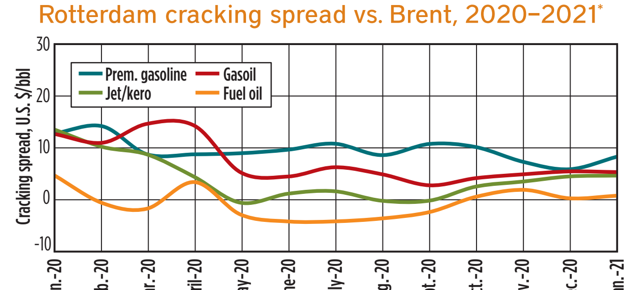

## Global refining utilization rates, 2020–2021*



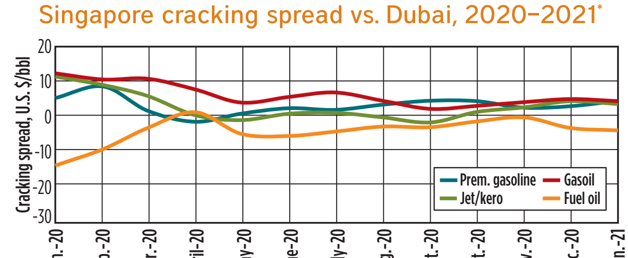

## U.S. Gulf cracking spread vs. WTI, 2020–2021*




## Rotterdam cracking spread vs. Brent, 2020–2021*




## Singapore cracking spread vs. Dubai, 2020–2021*

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# Finding upgraded sealing solutions

By universal agreement, the safety and reliability of facilities in the hydrocarbon processing industry—and most other industrial plants—are of highest importance. That is why modern facilities no longer consider mechanical seals as commodity products wherein only initial cost and timely delivery are given emphasis.

At reliability-focused facilities, mechanical seals are considered engineered products. Best-in-class companies involve one or more of their reliability technicians or engineers in the selection of preferred seal manufacturers. As the quasi-"owners" of a portion of the plant's machinery, they are intimately involved in vendor selection and performance tracking of mechanical seals. Knowledge of best practices and the advocacy of long-term sealing solutions are listed on the role statements of reliability professionals at best-in-class companies. An actual field example highlights how senior reliability technicians and reliability professionals contribute to mechanical seal success.

When the failure frequencies of sealing products in light hydrocarbon services at their plant did not meet expectations, two professional employees made thorough experience checks. After establishing that Plan 53 was very successful in comparable services elsewhere, they presented their firm recommendations to management. They reached consensus that Flush Plan 53, incorporating a heat exchanger in the flush loop (**FIG. 1**), should be recommended for services up to 430°F (220°C) at their U.S. refinery.

However, close communication with one of the plant's technology providers indicated that for several pumps operating with temperatures substantially above 400°F (205°C), the heat load in a Plan 53 circuit could exceed comfortable factors of safety. Attention was then concentrated on API Flush Plan 54 (**FIG. 2**), where an external pump is used to circulate flush fluid in a closed loop. Plan 54 offers a high rate of heat dissipation and positive circulation of flush fluid. For the record: If properly maintained, this is often considered the most reliable pressurized plan for dual seals. HP

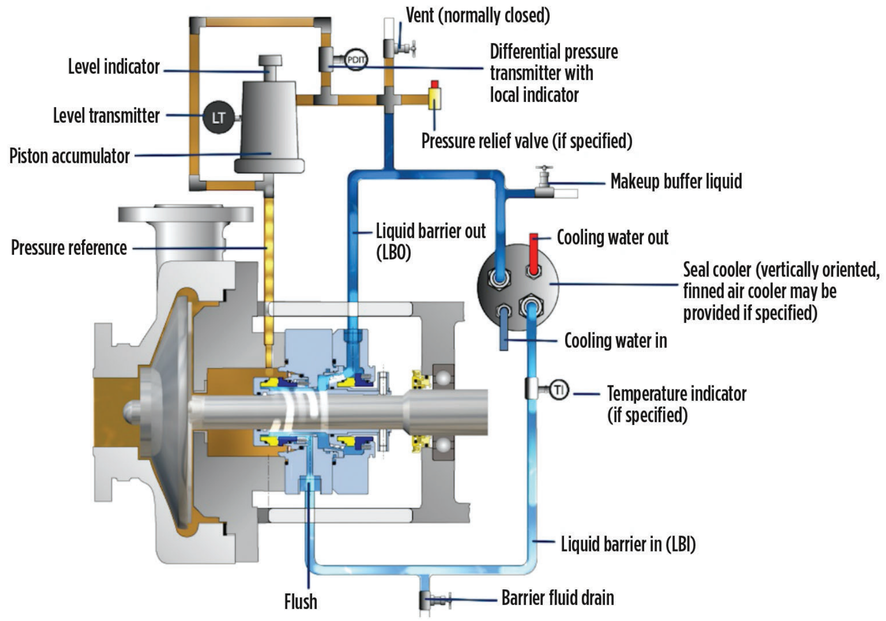


**FIG. 1.** API Plan 53C: Pressurized and cooled barrier fluid circulation in the outboard seal of a dual-seal configuration. A tapered pumping ring keeps up circulation while running. Source: AESSEAL Inc.

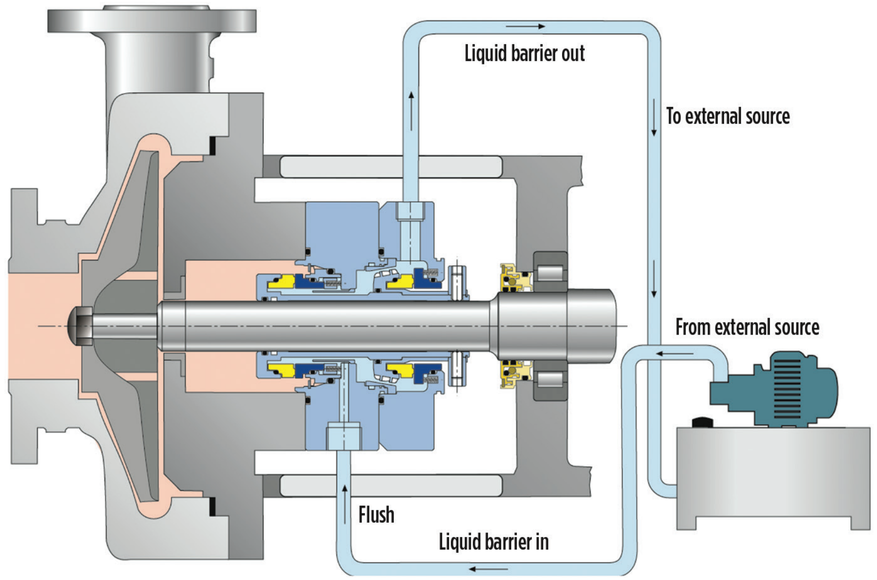


**FIG. 2.** API Flush Plan 54. Source: AESSEAL Inc.

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# Case 111: Predict the cause of cracks in gear teeth and time to failure

In the practice of engineering, we sometimes wish for better hindsight or foresight. This may occur when management questions why something has occurred, and what the risk is to operate until the next planned turnaround. In such a scenario, management is essentially asking for a prediction of the past *and* the future.

Consider the following case history: A crack is noticed on a gear tooth after a sudden power outage. The risk is that if the tooth breaks off and falls into the gear mesh, a major wreck could occur, resulting in a long outage. Knowing when and why the crack started is also important to ensure that a repeat event will not occur. Starting up after a gear set replacement without knowing the cause of the crack is always risky.

Crack growth-type equations can be used on ductile steels. Knowing or assuming an initial crack size and knowing the cyclic stress trying to open the crack, the cycles to grow the crack to a certain length can be estimated.

For this case, it is assumed that some event started a crack in the tooth root across the tooth face, as shown on **FIG. 1.** It takes a very large shock load of many times the allowable stress to cause a crack across the face of a case-hardened gear tooth.[1] The case depth may be $a_i$ = 0.05 in. deep in the tooth face width ($b$). When the crack was first noticed, it appeared to have grown to $a_f$ = 0.13 in. length. When the crack progresses halfway through the tooth as shown, it will break off in a bending-type failure at $a_f$ = 0.5 in. This happens because as the crack grows, it reduces the support material.

The cycles ($N$) to grow the crack can be approximated by integrating the Paris Differential Equation, shown in Eq. 1:

$$da \div dN = C \times (\Delta K)^m \qquad (1)$$

where:

$dN$ = Cycle length
$da$ = Crack growth during the cycle
$\Delta K$ = Stress intensity at the crack tip
$C, m$ = Material constants.[2]

For a large plate where the cross-section does not significantly change as the crack grows, the constant surface bending stress ($\Delta\sigma$) from zero to peak stress can be used, as shown in Eq. 2:

$$N = \left[8.3 \times 10^8 \div \Delta\sigma^{3.25}\right] \times \left[1 \div a_i^{0.625} - 1 \div a_f^{0.625}\right] \qquad (2)$$

The calculation is different for a rectangular beam-shaped gear tooth, as shown in **FIG. 2.** The stress becomes an iterative procedure at each growth step, since the bending resistance of the supporting material decreases as the crack grows. Only two growth steps are used in this example, with the average stress in each interval labeled $a$. **TABLE 1** shows the results from the calculations.

From this data, it appears that the power outage could have been the cause, since sudden power outages can result in a large gear tooth shock load.[3] Once started, the crack grows very quickly. Caution suggests not starting up the unit without a gear set replacement. This is an approximate method of looking into the past and future, but it does help lower the risk in making a decision. HP

**TABLE 1.** Crack growth with stress

| Cyclic stress on tooth | $a_i$ | $a_f$ | $N$ | Life at 1,200 rpm |
|---|---|---|---|---|
| $\Delta\sigma$ = 11.6 ksi to start<br>$a$ = 0.09 in. average | 0.05 in. | 0.13 in. | $8.7 \times 10^5$ cycles | 12 hr ago |
| $\Delta\sigma$ = 16.9 ksi to fail<br>$a$ = 0.31 in. average | 0.13 in. | 0.50 in. | $1.8 \times 10^5$ cycles | 2.5 hr to fail |

**NOTE**

*Case 110* was published in *HP* November 2020. For past cases, please visit www.HydrocarbonProcessing.com.

**LITERATURE CITED**

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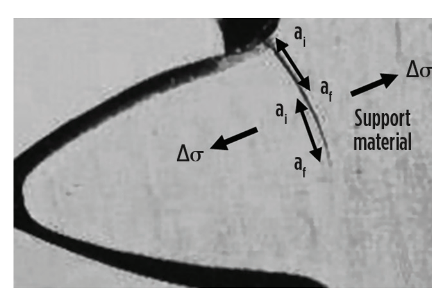
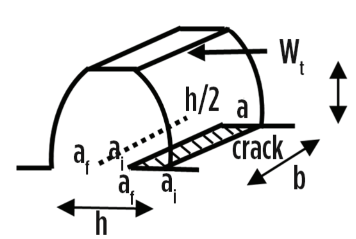



**FIG. 1.** Crack in a normal gear tooth.




$\Delta\sigma_{surface} = 6 \times W_t \times L/(b \times h^2)/1{,}000$ ksi
$\Delta\sigma_{crack} = 6 \times W_t \times L/(b \times (h - a)^2/1{,}000$ ksi

**FIG. 2.** Crack in a rectangular beam-shaped gear tooth.

HP Executive Viewpoint

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# Interconnected neural networks drive breakthrough optimization

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We have all learned that technology alone will not improve our operations, it is how we use technology as an organization and align our teams around it. Using artificial intelligence (AI) to improve the continuous optimization of a refinery or chemical plant is no different. Augmenting your current optimization methodologies and procedures with AI will not create a step-change in plant process optimization because the AI is limited by convoluted team structures and siloed modeling tools.

**Limitations of the traditional control and optimization hierarchy.** The sheer complexity of controlling and optimizing plants gave rise to what we recognize today as the traditional hierarchy of technologies and teams. Under this hierarchy, a narrow top layer—planning and economics—governs the lower layers. Below this planning layer is process engineering, which is responsible for operating areas and units in the plant in accordance with the targets provided by the upper layer. Below the process engineering layer is process control. Oftentimes, this layer is divided into two parts: advanced process control (APC), which includes multivariate model predictive controllers, and a distributed control system (DCS), which regulates valves via proportional integrative derivative (PID) controllers. The bottom layer of the hierarchy is the field instrumentation layer of valves and indicators.

This hierarchy is designed such that the most economically critical decisions are made at the top, typically on a weekly basis. The decisions then cascade down through the layers, eventually controlling on a minute-wise or second-wise basis thousands of process variables in the form of pressure, temperature, flow, level and quality specifications. Process engineering and operations teams are familiar with carefully managing and optimizing these thousands of variables. Keeping each of these variables optimized at all times is collectively important to plant profitability. However, most of them are not individually critical to keep optimized continuously. By contrast, hidden amongst these thousands of variables are 10–15 pivotal process variables that have asymmetric economic importance. If these 10–15 process variables are identified and optimized continuously, refineries could generate \$20 MM–\$30 MM in additional margin from their most profitable processes.

One example of a pivotal process variable can be found in conversion units. Planning and economics will provide a general target for conversion or reactor temperature, accounting for the feedstock and the product economics. What if the reactor could always run at the optimal temperature, down to 1° of accuracy every hour of every day? What if this temperature continuously adapted to small unmeasured changes in feed composition and unit conditions to avoid under-cracking or over-cracking the molecules? The answer is millions of dollars of incremental annual value in the product pool.

Under the traditional control and optimization hierarchy, the plant has no way of keeping such a conversion variable tightly optimized at all times, because it requires making decisions at a plant-wide scope. Plant-wide decisions are made at the top planning and economic layer on a weekly basis. By the time the economic decision trickles down to the process control layer, the unmeasured feed disturbance will be long gone and the opportunity to convert the feed more accurately is lost.

The conventional hierarchical control and optimization layered structure is designed for "regular" variables, where general economic guidance is satisfactory, and the precise control of the variable can be done at the local level. The conventional hierarchy is not positioned to keep these "special" variables optimized continuously, since this combines the local process scope and the global plant scope. Unfortunately, the current trend of integrating AI into the different layers of the hierarchy can provide some local value, but not fix the fundamental structural problem. This is where the processing industry has a breakthrough opportunity for a process optimization step change.

**Thinking beyond the traditional control and optimization hierarchy.** Capturing the opportunity behind

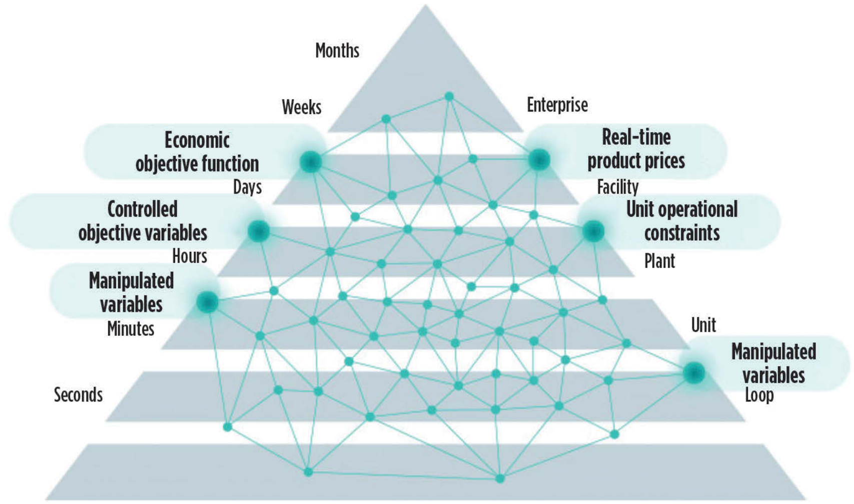


**FIG. 1.** The interconnected closed-loop neural network integrates operations, processes and variables throughout the traditional hierarchy.

the asymmetrically important special process variables requires us to change our mental model around the hierarchy of control and optimization. Similarly, operating the thousands of regular variables must be kept intact, along with the layers and teams that control them. However, we must create an interconnection between the special process variables, related process variables in other parts of the plant and higher layers such as process engineering and planning and economics.

This interconnection is not only between models at the different layers of the control and optimization hierarchy, but between people in the different layers. This interconnection should take the form of a single model that aligns the different teams from various disciplines and from different areas of the plant, all centered around the continuous optimization of the special variables. This should all be done side-by-side and without any disruption to the traditional control and optimization hierarchy, keeping the thousands of regular variables intact.

**An interconnected neural network to optimize key plant variables.** This interconnection is already happening in multiple plants across the industry. It is empowered by neural networks that have revolutionized the way refiners approach and manage optimization for their most complex and profitable processes (**FIG. 1**). These neural networks can identify the 10–15 special variables that, if optimized continuously, can have the biggest economic impact. These neural networks align people across different teams on the most valuable problems and operate in closed loop to unlock hidden margins across multiple plant process units.

One further substantial advantage of the interconnecting neural network is the ability to compensate for the lack of reliable real-time feedstock compositional data. The neural network can compensate for this typical lack of composition data using pressure, temperature, flow and level indicators that react to feed composition changes through patterns that are discovered by the neural network. Using historical data, the neural network learns subtle nonlinear relationships between the special variables and manipulates them either through APC targets or PID setpoints.

**Takeaway.** The traditional control and optimization hierarchy, combined with advancements in software technology and tools, has supported increasingly complex operations over the years. However, we are all looking for ways to operate better and rethinking how we are organized to run the plant more optimally. It is imperative to understand the asymmetrical criticality of the continuous optimization of a very small group of special process variables. Through the integration of a closed-loop neural network, refineries and petrochemical plants can interconnect their people and process areas to optimize their most dynamic and profitable processes and capture millions of dollars in annual incremental margin. HP

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# Russia looks to double petrochemicals and LPG exports by 2025

Russia hopes to become one of the world's largest producers and exporters of petrochemicals and LPG by 2025 by doubling its domestic output and exports. Implementation of this plan will be part of the country's existing International Cooperation and Export initiative, providing a roadmap for the expansion of the petrochemicals and gas processing sectors. Despite Russia's status of one of the world's largest oil and gas producers, its share of petrochemicals and LPG in the overall structure of the Russian economy does not exceed 2%.

**Promoting petrochemicals and LPG.** As part of its development plan, Russia aims to increase the value of its petrochemical exports to $37 B by 2024 from $22.4 B in 2019. To this end, domestic petrochemicals production will be scaled up rapidly over the next 4 yr–5 yr.

For example, the production of large-scale polymers should reach 11.1 metric MMtpy by 2025 from 5.3 metric MMtpy in 2019, while exports of these polymers will increase to nearly 4.4 metric MMtpy from 600,000 metric tpy. Simultaneously, the share of LPG processing in the country is planned to increase to 8.2% from the current share of 4.6%, while naphtha processing will rise from 5.6% to 7.2% by 2025.

In Russia, gas has been traditionally used as an energy source to generate electricity and heat, rather than as a valuable feedstock. That mindset is now changing. Analysts believe that NGL production is set to gain momentum in Russia. The opportunity is certainly abundant—according to the Russian Ministry of Energy, only 12% of the overall volume of gas produced in Russia in 2019 was utilized for further processing.

**Tax breaks for producers.** The provision of tax breaks for producers, particularly a reverse excise tax on ethane and LPG, is expected to stimulate interest in creating a larger export economy around petrochemicals and LPG. Several leading Russian energy firms have been calling for such tax breaks for some time.

According to the Russian Ministry of Energy, the provision of a reverse excise tax on ethane and LPG will attract more than RUB 3.5 T ($47.7 B) of additional investment in Russian petrochemicals and gas chemicals over the next 6 yr–7 yr, as well as increase the processing of ethane and LPG by 8 metric MMtpy–10 metric MMtpy.

**Companies respond.** Due to infrastructure constraints, ethane processing in Russia is severely under-developed. According to a report in Russia's *Vedomosti* newspaper, Russian gas contains approximately 10 metric MMt–12 metric MMt of ethane, of which only 700,000 metric t are further processed. The volume of LPG production in Russia is estimated at 17 metric MMtpy, of which 40% is exported.

The announcement of the petrochemicals and LPG export development plan may also help stimulate companies to resume work on their investment projects in Russia. Many of these projects were suspended due to the COVID-19 pandemic and its negative economic consequences.

A number of leading local producers have already announced their intention to look into new investment opportunities in Russia's petrochemicals and gas chemicals sectors. One company is Lukoil, a privately owned oil and natural gas producer and oil refiner (FIG. 1). Lukoil has confirmed plans to build new capacities for the production of polypropylene (500,000 metric tpy) and styrene (up to 300,000 metric tpy) at its refinery in Nizhny Novgorod in Russia's Volga region. The petrochemicals complex will be a new project for the company.

**FIG. 1.** Lukoil operates four oil refineries in Russia: Perm, Volgograd, Nizhny Novgorod and Ukhta. Photo courtesy of Lukoil.

Lukoil is also considering the construction of a large-scale petrochemical complex at its Perm refinery. The cost of the project is estimated at more than RUB 200 B ($2.73 B), which would make it one of the most expensive projects in the Russian petrochemicals sector.

**Arctic gas processing developments.** As part of Russia's export plan, a development has appeared in the Arctic region. To ensure 80 metric MMt of cargo turnover along the Northern Sea Route by 2024, the Russian Ministry of Economic Development has approved a package of incentives for the gas chemicals sector.

Furthermore, Lukoil intends to invest RUB 610 B ($8.3 B) in a gas processing complex off the coast of the Gulf of Ob for the production of methanol, ethane and LPG. Meanwhile, Gazprom Neft has vowed to allocate up to RUB 1.1 T ($15 B) in the production of ethanol, polyethylene and polypropylene production within the territory of the region.

Russian producers welcome the government's plans to financially support the domestic petrochemicals and gas processing sectors. The present volatility of prices for finished products and raw materials poses a threat to payback periods for many of large-scale petrochemicals and gas processing projects, making the long-term profitability of such projects difficult to predict. HP

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# ROI is the secret sauce for sustainable digital transformation

Digital transformation is a broad term that can mean vastly different things to different people. Within a given company, it is easy to gain consensus that such a thing is necessary to stay competitive, but it is tremendously difficult to define what it looks like in practical terms—much less where to begin.

Of course, it is infeasible to lay out a sequence of steps that is right for every company in every industry. Rather, it is more useful to talk about digital transformation in terms of levels of maturity, each centered around the key variable needed to make the process sustainable: return on investment (ROI). For digital transformation to be beneficial and long-lasting, it has to complement business goals. It is important to keep expectations realistic with respect to the current level of digital maturity. Five such levels of digital maturity are discussed here.

**The experimental stage.** An organization in the experimental stage has acquired some basic digital tools but still lacks a cohesive vision for what it hopes to achieve. Solutions built during this stage are commonly built by intellectually-curious individuals attempting to solve small-scale problems. These efforts can provide an ROI that is well-aligned to business priorities within this individual's area of responsibility, but these contributions are unlikely to be easily scalable or repeatable.

**The aggregation stage.** Organizations in this stage are recognizable by their ongoing efforts to collect increasingly more data in one place. Some may narrow their focus to a specific target area of the business so they can quickly move on to the next phase of this maturity model within this scope. Meanwhile, others will attempt to collect as much data as part of a broader scope. Either way, for ROI to be realized at this stage, discipline is critical. It is important to try and add context to the collected data, where possible. This extra information will help ensure it can be effectively wielded by as large a group as possible.

**The KPI stage.** Well-designed key performance indicators (KPIs) reflect how well something is doing within a given scope (i.e., enterprise, division, area or unit). An organization in the KPI stage would hopefully leverage them in two ways. First, these KPIs should guide both short-term and long-term strategy. ROI here comes from strategic decisions being backed by evidence. Secondly, by setting tolerable ranges for KPIs, the business can be alerted when a particular group may be falling behind and needs assistance. The ROI is realized here by ensuring everyone in the organization is able to play their part in the overall business strategy and is getting the help they need.

**The prediction stage.** Once enough data is available to form KPIs that reflect how individual groups are contributing towards overall business goals, organizations in the prediction stage are in a good position to forecast what state a system will be in and when. This impacts ROI in two ways. The first lies in the triggering of interventions in the process—it is possible to carry these out pre-emptively if a downward trend with respect to an important KPI is detected early enough (whether this means mobilizing resources or changing strategy). The second impact revolves around the refinement of this strategy. More robust models of the interplay between variables means evidence can be more effectively wielded to achieve the intended outcome.

**The actuation stage.** Unlike the preceding levels of digital maturity, organizations in the actuation stage are much more able to start using digital tools not just for *knowing*, but for *doing*. This means an organization's prediction capabilities have earned so much confidence that digital tools no longer just inform decisions but are able to carry them out in real time. An excellent example of this is an advanced process controller (APC). Instead of people having to specify set points for their control systems to hit, tools at this stage can calculate a set point based on a set of optimization criteria for the process. The ROI seen at this stage can be quite lucrative. Not only does automation allow for a much finer level of control that can more rigorously optimize for business-critical output variables, but it frees up human operators' attention for other things.

**Takeaway.** Needless to say, advancing through these stages can be quite an undertaking. Again, ROI should be a key variable for prompting leaders on where to focus. Having a good grasp of the ROI in a specific situation provides an idea of what resources can be viably committed, as well as what expectations should be along the way. Digital transformation is a journey and being able to promise and deliver on ROI is key to sustaining it over the long term.

However, the skills needed for such an effort can be hard to find. Fortunately, there is a rich ecosystem of vendors and service providers that offer the tools and the expertise needed to develop a transformation plan, put it into action and keep it on track. Even if the intent is to build these capabilities in house, there may be a clear ROI to supplementing teams with expertise from outside. HP

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# Fuels and chemicals: Finding the right refinery configuration for a less predictable world

The refining industry links the upstream production of crude oil with the end markets for fuel products, as well as for the petrochemical/chemical industry. Refineries have been investing in complex facilities to improve conversion and to better alter the final product composition and quality to adhere to changing market and regulatory requirements. For example, by using a combination of or stand-alone fluidized catalytic cracking unit (FCCU) and hydrocracking capacities, refineries can orient output to adhere to a wide range of gasoline/diesel ratios. Similarly, a bottoms upgradation facility enables refiners to eliminate or minimize fuel oil per market demands. However, a mismatch in refinery output and local market demand can still exist. This predicament drives the need to either export fuels or divert product volumes to petrochemicals/chemicals. Drivers for higher petrochemical feedstocks were traditionally regional and driven by a combination of the local fuel demand mix (e.g., diesel/gasoline ratio and LPG demand/composition), competitiveness in fuel export markets and the availability of natural gas as feedstock for steam crackers.

Petrochemicals are projected to be the largest driver of world oil demand growth, surpassing that of gasoline or diesel by 2030. Drivers for this change are the growing electrification of vehicles and the increasing fuel efficiency of new vehicles. Therefore, integrating petrochemical production capacity into refineries is imperative to solve the mismatch in local market supply and demand. In this scenario, non-integrated refiners and petrochemical producers with less flexibility will be more vulnerable to demand risks. Refiners and chemical producers will respond by developing integration synergies uniquely tuned for each site. However, there is no one-size-fits-all option. The best option will be chosen by examining product opportunities, facility configurations, technologies, logistics and return on capital.

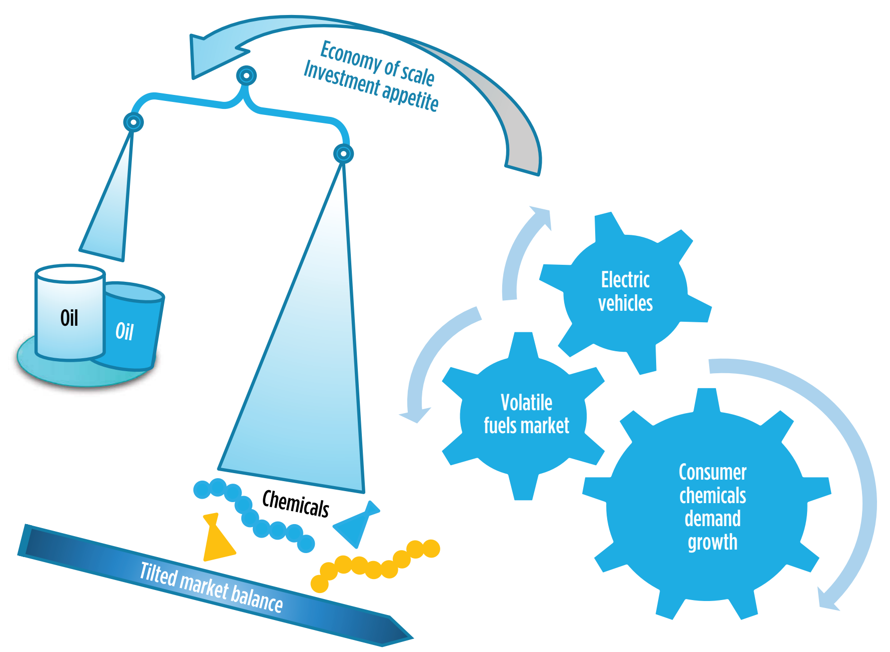


**FIG. 1.** Thought diagram showing that future projections of consumer chemicals demand are stronger than projections of fuels demand.

The most common and practicable ways that refineries are integrated with petrochemical production include:

- A refinery integrated with a steam cracker
- A refinery integrated with an aromatics complex
- A refinery integrated with a steam cracker and aromatics complex
- Downstream chemicals integration with the above routes.

In conventional refinery integration, excess naphtha and LPG were allocated to the cracker and aromatics complex. In the revised scenario of plateauing fuels demand, refiners have started selecting technologies with the intent to divert more distillates toward steam crackers and aromatics. Alternately, some refineries elect to focus on fuels and to divert excess naphtha/LPG to a separate petrochemical facility rather than having an integrated complex. These are two extremes of a continuum of possible refinery configurations.

## Synergies: Stand-alone vs. integrated refinery/petrochemical complexes.

In the non-integrated scenario, refineries have been selling naphtha to other independent steam cracker and aromatic complexes. However, there are significant premiums in integrating these complexes at a

single site, which allows consolidation of the intermediates and monetization of the synergies. The advantages are multifold:

- Olefins command significant premium over transportation fuels
- The impact of demand and price fluctuations on profitability are dampened
- More options exist to alter product slates to respond to market needs
- Ensured feedstocks availability for petrochemicals
- Value enhancement from integrating intermediate streams
- Capital and operating costs and resource optimization, savings from which can be further attributed to the following:
    - Shared infrastructure and utilities
    - Lower transportation costs
    - Minimization of fixed overheads.

Some of the strategies that can be critical in the economics of integration are discussed in the appending section. These include:

- Hydrogen sharing between the refinery, cracker and aromatics increases flexibility
- Combining ethylene/propylene recovery from the FCCU/coker offgas with the cracker
- Multiple options in optimizing the $C_4$ stream, including producing butadiene/butenes, hydrogenation and recycle to the cracker, or adding to LPG
- Upgrading pyrolysis gasoline to avoid the production of negative value products (i.e., those with value lower than the cracker feed). Some options include:
    - $C_5$ diolefins can be utilized to manufacture dicyclopentadiene, depending on market requirements. Alternately, $C_5$s can be hydrogenated and recycled to the cracker or added to the gasoline pool.
    - Pyrolysis gasoline contains over 60 wt% $C_6$–$C_{10}$ aromatics, half of which is benzene. The $C_6$ cut can be fed to extractive distillation for producing benzene.
    - The $C_7$+ cut has an excellent octane rating and can be advantageously sent to the gasoline pool. When demand for benzene is high, the $C_7$+ cut can be sent to the hydrodealkylation process to be converted to benzene.
    - The $C_7$–$C_8$ streams can also be integrated with the aromatics section for paraxylene (PX) yield enhancement.
    - The $C_9$+ cut can be absorbed in the diesel pool.
- Raffinate from the aromatics complex is blended in the gasoline pool.
- Heavy aromatics from the aromatics complex is blended in the diesel pool.

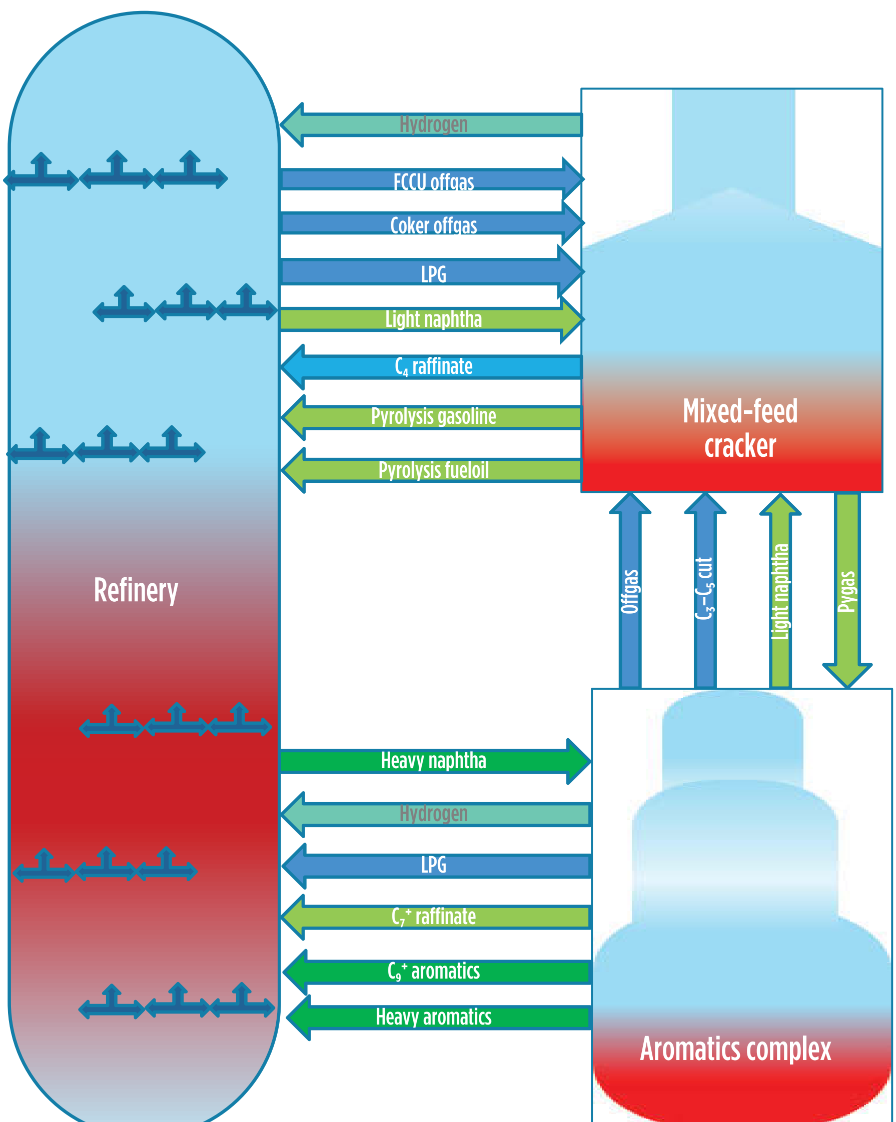


**FIG. 2.** Synergies among refinery, petrochemical and aromatics complexes.

**FIG. 2.** illustrates the common synergies between the refinery, petrochemical and aromatics complexes.

A high-level integration case study has been developed to illustrate the possible alternate grassroot configurations and their directional impact on refinery margins and rates of return. The location of the complex is a key driver in overall project economics, since relative fuels/chemicals pricing, fuel costs, crude/product freight impact and capital expenditures (CAPEX) are all dependent on where the project is located. For the purpose of this study, the location of the new complex is in India.

**Fuels and petrochemicals outlook for India.** India is emerging as one of the fastest growing economies in the world. With a total capacity of 250 MMtpy, India is the fourth largest refiner in the world. It is positioned to serve domestic demand of petroleum products and to supply petroleum products to other Asian countries.

A high population and relatively low per capita consumption show the promise of high economic growth rates for the energy sector and consumables. In addition to fuels, chemicals and petrochemicals also have vital stakes in economic growth—namely agriculture, infrastructure, health care, textiles and consumer durables. Petrochemical products cover the entire range of consumables and durables, ranging from clothing, housing, construction, furniture, automobiles, household items, toys, agriculture, horticulture, irrigation and packaging to medical supplies. Between 2015 and 2040, India's GDP growth rate is forecast to be more than 7%.

Some relevant projections for the Indian refining and petrochemicals sector are:

- Petroleum product demand is forecast to increase to 472 MMtpy by 2040.
- Diesel and gasoline consumption are forecast to increase at a compound annual growth rate (CAGR) of more than 4% to 2040.
- Domestic refining capacity is projected to increase by 224 MMtpy by 2040.
- More than 5 MMtpy of polypropylene (PP) capacity is expected to be added by 2025.
- More than 4 MMtpy of high-density polyethylene (HDPE) and low-linear-density polyethylene (LLDPE) capacity is forecast to be added by 2025.
- By 2025, India is forecast to have a deficit of purified terephthalic acid and PX of 1.8 MMtpy and 600,000 tpy, respectively.

Additional parameters relevant to the Indian fuels and chemicals scenario that have a significant bearing in setting the boundaries of the integration case study are as follows:

- Fuel demand is skewed toward diesel (3:1 diesel-to-gasoline ratio).
- Naphtha is the primary feedstock for steam crackers, since cheap domestic natural gas is scarce.
- LPG specifications can absorb up to 80% $C_4$s.
- Research octane number (RON) and motor octane number (MON) specifications for regular grade Bharat Stage 6 (BS-6) gasoline (91 and 81) are lower than Euro 6 gasoline (95 and 85).
- Olefin content is limited to 21% in the BS-6 specification for regular grade gasoline. This is slightly greater than the maximum of 18% specified for Euro 6 gasoline.
- Capital costs in India are significantly lower vs. in the U.S. Gulf Coast and the Middle East.
- A significant domestic supply shortfall (more than 50% by volume) in several petrochemical intermediates exists, primarily from six major value chains in petrochemical intermediates.
- Utility costs are higher, as utilities are majorly captive, and centralized utility providers are rare.

**Case study methodology.** The case study illustrates the investment analysis for a hypothetical integrated refinery/petrochemical complex. The objective is to demonstrate the impact on refinery margins and returns as the product slate is shifted from fuels to petrochemicals. A simple model is constructed to reasonably capture the product slate of the alternate configurations based on typical yields from different technologies. This simplified model has constant yields in units (such as FCCUs) in high-severity mode with a fixed feed of hydrotreated vacuum gasoil (VGO), and does not vary, thus optimizing the feed, operating conditions or yields for each case. There are tradeoffs in selecting this simplified approach in relation to the rigorous LP model. This model will not simulate all possible combinations of design variables and optimized solutions. Uncertainty exists in future demand projections, and directly coupling of such highly estimated data with sophisticated LP subroutines will not add value. Alternately, the results from this directional assessment can be utilized as guides in screening and for carrying out site-specific optimizations using the LP model. Feedstock and product prices are based on 3-yr averages, which should be adequate to capture average refinery margins (FIG. 3). The order-of-magnitude CAPEX for calculating the internal rate of return (IRR) is based on adjusting/scaling publicly available estimated project costs for similar projects/units in recent years.

**Base case.** The base case (FIG. 4) refinery is configured for processing 15 MMtpy of light crude (API > 33), producing

gasoline and diesel fuels. The base case refinery is equipped with a high-severity FCCU as the main VGO conversion unit, with an upstream VGO and heavy coker gasoil (HCGO) hydrotreater. The delayed coking unit (DCU) is selected for vacuum residue conversion. In line with demand patterns, the crude unit maximizes diesel production by dropping heavy-end kerosene material into the diesel cut. This configuration is capable of maximizing diesel, while still retaining the flexibility to add on petrochemical blocks by changing crude unit cut points and redirecting streams. Propylene is included as one of the intended major products from the base case refinery, which includes two 400,000-tpy PP units. The FCCU and DCU have world-scale capacities of more than 4 MMtpy.

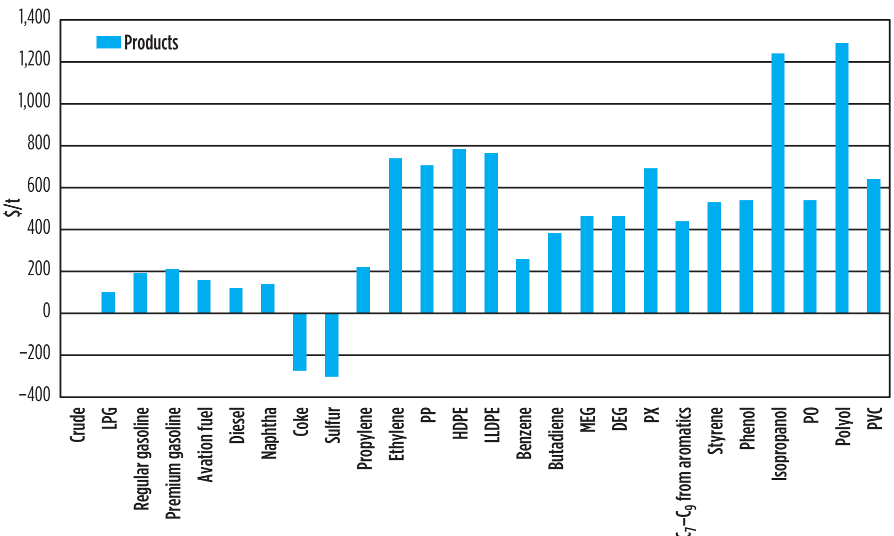


**FIG. 3.** Price set—Added value on crude.

**TABLE 1. Unit-level definition of the integrated case studies**

Crude capacity = 15 MMtpy

| Configuration | Base Case | Case 1 | Case 2 | Case 3 | Case 4 |
|---|---|---|---|---|---|
| **Refinery units** | | | | | |
| CDU/VDU | X | X | X | X | X |
| DCU | X | X | X | X | X |
| VGO-HDT | X | X | X | X | X |
| FCCU | X | X | X | X | X |
| Kerosene hydrotreater | X | X | X | X | X |
| Gasoline hydrotreater | X | X | X | X | X |
| CCR | X | X | | X | X |
| NHT | X | X | X | X | X |
| Isomerization | X | X | X | X | X |
| ERU | | X | X | X | X |
| **Petrochemical units** | | | | | |
| Aromatics complex | | X | | X | X |
| Mixed-feed cracker | | | X | X | X |
| HDPE/LLDPE | | | X | X | X |
| Ethylene oxide/ethylene glycol | | X | | X | X |
| Propylene oxide/styrene monomer (SM) | | | | | X |
| Polyol | | | | | X |
| PP | X | X | X | X | X |
| IPA | | | | | X |
| Cumene/phenol | | | | | X |
| Butadiene/1-butene | | | X | X | X |

## Integration cases.

The integration cases explore the common integration strategies and test the boundaries of the chemical conversion and economic margins that can be achieved. The following describes the integration case studies:

- Case 1: Refinery integrated with a steam cracker
- Case 2: Refinery integrated with an aromatics complex
- Case 3: Refinery integrated with a steam cracker and aromatics complex
- Case 4: Refinery integrated with a steam cracker and aromatics complex and downstream chemicals.

**TABLE 1** provides a unit-level definition of these integration cases.

## Case 1.

Case 1 adds an aromatics complex to produce 1.2 MMtpy of PX from naphtha, as well as an ethylene recovery unit from FCCU/DCU offgases to produce mono-ethylene glycol (MEG) to the base case refinery. The units in the aromatics complex are high-severity reforming (continuous catalytic reformer), PX separation and purification, $C_8$ raffinate isomerization, toluene and $C_9$–$C_{10}$ transalkylation, and benzene-toluene extractive distillation. The products produced from the aromatics complex are PX, benzene and hydrogen.

## Case 2.

Case 2 adds a 1.5-MMtpy mixed-feed cracker and downstream polymer units to the base case refinery. The key feedstocks considered for the cracker include:

- Offgases from the DCU (rich in ethane and olefins)
- Offgases from the FCCU (rich in ethylene and ethane, with some propane)
- LPG from the refinery pool
- Straight-run naphtha and kerosene streams to satisfy cracker capacity.

The proposed downstream configuration for Case 2 is:

- A world-scale swing unit to produce HDPE and LLDPE
- A standalone HDPE unit
- A 1-butene facility to satisfy the requirement in PE plants as a co-monomer
- Additional chains of PP units to convert propylene produced from the complex
- The generated $C_6$ stream from pyrolysis gasoline hydrogenation will be charged into the benzene extraction unit from where $C_6$ raffinate is recycled back to the cracker
- The hydrogenated $C_5$ cut is recycled back to the cracker
- The $C_7$+ stream is routed to the gasoline pool.

MEG production—diethylene glycol (DEG) and triethylene glycol (TEG) as byproducts—is not considered in this case, and all ethylene produced is consumed in the HDPE/LLDPE units.

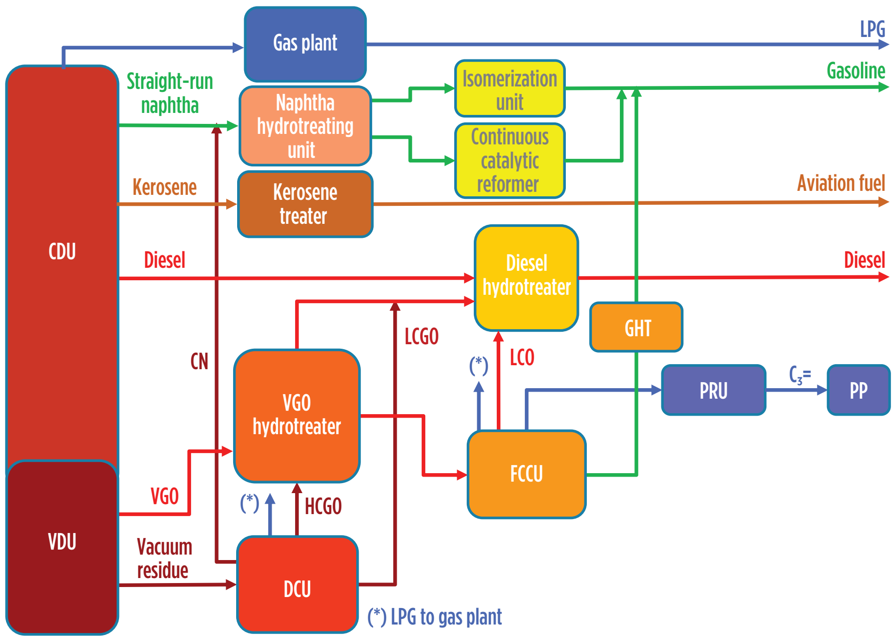


**FIG. 4.** Base case: Fuels refinery.

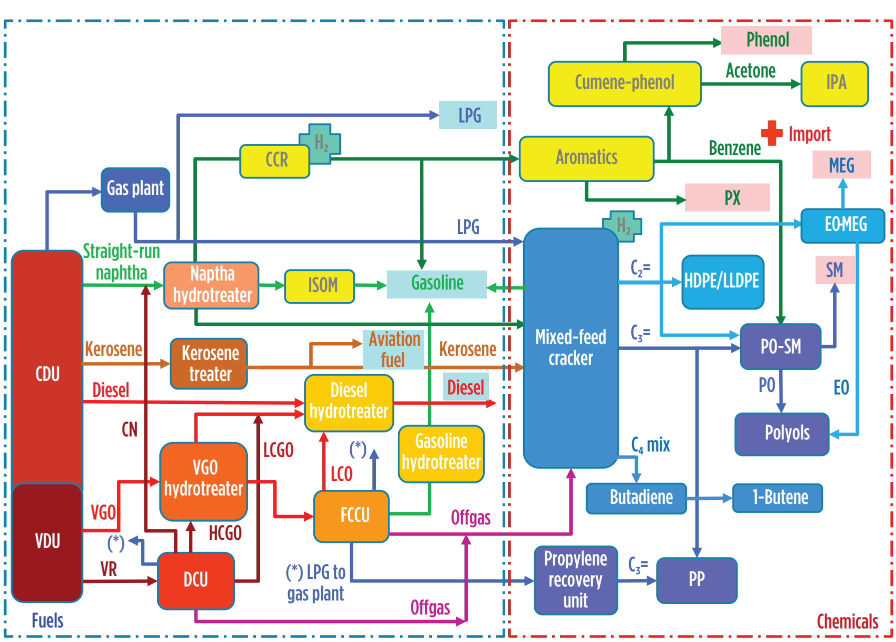


**FIG. 5.** Case 4 block flow diagram: Refinery, petrochemical and aromatics complexes, and downstream chemicals.

**Case 3.** Case 3 combines Case 1 and Case 2 with the base case refinery (i.e., the aromatics complex and mixed-feed cracker in one single configuration). In addition to the standard downstream petrochemical units present in Case 2 (namely HDPE, LLDPE, PP and butadiene/1-butene), Case 3 also includes MEG production.

**Case 4.** This case is a modification of Case 3, adding downstream chemicals value chains for propylene and ethylene. These include cumene/phenol/isopropyl alcohol (IPA), propylene oxide (PO)/propylene glycol/polyol and styrene (as a co-product of PO). A significant demand and growth rate exist for these downstream specialty petrochemicals to support an investment in India. Case 4 is the highest extent of chemical integration that this case study explores. **FIG. 5** represents a high-level block flow diagram to detail the major streams.

Considering local market requirements for a higher diesel-to-gasoline ratio, each configuration ensures that this ratio remains above 2, even as petrochemicals content is increased. In addition, LPG production is curtailed and diverted to the cracker in view of expected cheaper alternative sources. Diesel is the predominant fuel in India's fuel mix, and, with a significant forecasted increase in the adoption of electric vehicles, future gasoline demand is less predictable. The configuration achieves world-class, single-train capacities for all major units/blocks, while limiting the refinery capacity to 15 MMtpy. This helps limit capital investment, while avoiding penalties due to economies of scale. This configuration allows the focus to be on satisfying local/national market requirements, as well.

**Summary of results.** The product slates, refinery margins, relative CAPEX and IRR for all cases are summarized in **TABLE 2.** This also provides an idea of the multitude of capacities that can be achieved for each product chain.

**Sensitivity analysis.** The global refining and petrochemical industries are forecast to undergo a massive shift due to forces emanating from either end markets or upstream feedstock prices. The choice

of feedstock, demand patterns, government directives and return on capital requirements will influence configuration decisions. The following parameters of each case study have been varied to see the impact on each configuration:

- Lower diesel prices due to excess production
- Lower gasoline prices due to lower demand growth
- Lower prices for ethylene-based chemicals
- Lower prices for propylene-based chemicals
- Increase in CAPEX by 10%.

**FIG. 6** details the IRR for each case for these sensitivity scenarios.

**Summary of findings.** Even this limited high-level study utilizing a simplified methodology leads to interesting conclusions. Increasing the share of petrochemicals in the product slate increases refinery margins and IRR. In this case study, while growth in refinery margins is almost linear, returns start to plateau as petrochemicals content approaches 30%. The sensitivity analysis demonstrates that an integrated refinery is better able to manage returns, even with a decrease in cracks and spreads. This makes the integrated complex more robust and resistant to swings in demand. For a higher percent of crude-to-chemicals conversion to be economically viable, margins need to be harvested from crude pricing, export-oriented product pricing, and elaborate residue upgrade investments for heavier crude processing, among others.

Diverse product mixes, fuels and chemicals reduces risks from anticipated market disruptions. Harnessing value from intermediates, achieving economies of scale, limiting CAPEX, spreading fixed costs, accounting for local market drivers and logistics are key considerations in selecting the optimum configuration for a future refinery.

Several additional factors that need to be considered while specifying the configuration include:

1. A site-specific combination of factors (e.g., crude delivery infrastructure, location near demand markets, and land and construction costs) can boost or reduce refinery margins by \$4/bbl–\$8/bbl, which can significantly alter facility economics.
2. This case study considers a grassroots refinery focused on supplying the domestic market and exploiting local diesel demand. Integrating chemicals to an existing refinery or to a refinery focused on significant exports will be complex and will differ in several aspects.
3. Several options exist in configuring downstream chemicals. This case study selected a few of these options for illustration. For example, MEG is a capital-intensive process that is highly price sensitive. It has been considered in some cases, but not all. Another example is low-density polyethylene/ethylene vinyl acetate, a prospective niche petrochemical polymer, which is not considered here. This technology is licensed by very few licensors, as it has a high CAPEX and a significant proprietary content. Vinyl chloride monomer/polyvinyl chloride is also not considered, but can be added, provided an economical source for chlorine or ethylene dichloride can be found. Styrene and cumene/phenol/acetone have been included. This makes the benzene balance critical. A decision will need to be made on whether to import benzene or further convert the pygas $C_7$+ stream. For this study, importing benzene was considered. The value chain can be extended to include polycarbonates and

**TABLE 2.** Refinery margins and returns for alternate configurations

| Case ID | Base Case | Case 1 | Case 2 | Case 3 | Case 4 |
|---|---|---|---|---|---|
| Crude throughput, MMtpy | 15 | 15 | 15 | 15 | 15 |
| **Fuel products, thousand tpy** | | | | | |
| LPG | 1,275 | 1,275 | 600 | 600 | 600 |
| Gasoline | 3,050 | 2,100 | 2,100 | 1,900 | 1,900 |
| Aviation fuel | 500 | 410 | 300 | 200 | 200 |
| Diesel | 6,700 | 6,000 | 5,500 | 4,200 | 4,200 |
| **Total fuels, %** | **77** | **65** | **57** | **47** | **47** |
| **Petrochemical products, thousand tpy** | | | | | |
| PP | 750 | 750 | 1,350 | 1,350 | 820 |
| HDPE/LLDPE | 0 | 0 | 1,500 | 1,500 | 800 |
| MEG/DEG | 0 | 350 | 0 | 350 | 350 |
| PX | 0 | 1,200 | 0 | 1,200 | 1,200 |
| Benzene | 0 | 160 | 100 | 260 | (690) |
| Styrene | 0 | 0 | 0 | 0 | 750 |
| Butadiene | 0 | 0 | 70 | 70 | 70 |
| Phenol | 0 | 0 | 0 | 0 | 540 |
| IPA | 0 | 0 | 0 | 0 | 300 |
| Polyol | 0 | 0 | 0 | 0 | 400 |
| Petrochemicals, % | 5 | 16 | 20 | 31 | 31 |
| Gross refining margin, \$/bbl | 11 | 16 | 21 | 27 | 31 |
| Relative CAPEX at base case | 1 | 1.4 | 1.5 | 1.9 | 2.1 |
| **IRR, %** | **12.3** | **14.4** | **17.6** | **18.6** | **19.1** |

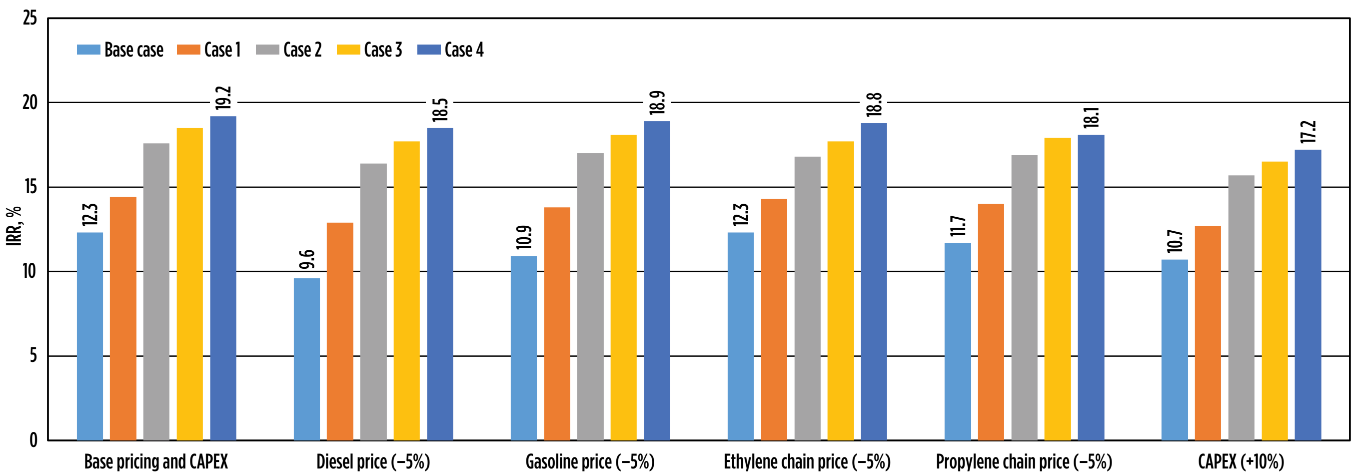


**FIG. 6.** IRRs for each case study for the sensitivity scenarios.

methyl methacrylates/poly methyl methacrylates. These can be taken up as site-specific enhancements.

4. The integration of petrochemical blocks increases CAPEX requirements and overall project complexity. This would require larger financial and execution resources. Construction time and ramp-up time until full production can be significantly higher for a large and complex integrated refinery, thus impacting return on investment (ROI).
5. The competitiveness of integrating petrochemicals with a refinery is dependent on two factors:
   a. Relative competitiveness of naphtha vs. cheap, natural-gas-based feedstocks for the cracker. Considering a low oil price regime, it is assumed that naphtha-based crackers will remain competitive in India.
   b. Integration is made more competitive by ensuring better integration with refinery streams and by achieving scale without diverting the diesel stream to the cracker or the aromatics complex. The investment costs are expected to be significantly higher in the latter option due to the additional units required to convert diesel to naphtha. Given the robust diesel requirement in the Indian market, this has not been considered in the current study.
6. The study is based on a single light crude; however, there does exist the option of utilizing heavier/cheaper crude slates. This will require changes to the refinery conversion configuration and alternate bottoms upgrading technologies, such as an ebullated-bed hydrocracker.
7. Several cutting-edge technologies are in the process of development and implementation. Thermal cracking of crude oil, along with improvements around the ethylene cracker to accept heavier feedstocks and higher-intensity hydrocracking, are some areas likely to attract significant research and development. When these newer technologies become commercially available, they have the potential to change the face of refining and petrochemical configurations.
8. Growth in consumer chemicals and plastics brings in environmental liabilities that are far too great to be taken lightly in the long term. Responsible growth strategies would require significant investments for the development of necessary infrastructures to support a sustainable culture of "protect, reuse and recycle."

Refining is a capital-intensive industry operating between two related, but independent, markets for crude oil and finished petroleum products. Configurations and operations that deliver adequate ROI are a function of a complex set of variables. Some variables, such as competition with alternate feedstocks and declining growth in transportation fuels, are global in nature. These require devising more diverse product slates by including petrochemicals. Other variables (such as local market dynamics, feedstock/product logistics and construction costs) are site specific and are to be used to arrive at the correct mix of fuels and petrochemicals that will remain profitable in different market scenarios. The case study presented provides directional insights into the paradigm shift from fuels to chemicals. The data presented should be treated as conceptual views of the authors and not directly attributed to the authors' organization. HP

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# How high can polyethylene and polypropylene plant capacities rise?

Polyethylene (PE) and polypropylene (PP) plants are the popular downstream derivatives of ethylene crackers or petrochemical fluidized catalytic cracking (petro-FCC). Refineries have been diversifying into petrochemicals to improve upon the gross refining margin (GRM) and flexibility in product mix. This has led to the evolution of the configuration of integrated refinery and petrochemicals complexes. Building such large complexes requires high CAPEX, so a rise in the demand to optimize cost economics is also seen.

Consequently, ethylene cracker plants with capacities exceeding 1 MMtpy of ethylene have become the norm. Ethylene plant capacities have soared close to 2 MMtpy, which was unthinkable two decades ago.

As the ever-growing thermoplastic, PE has been riding the wave and sharing the bulk of the ethylene pie. It is not uncommon to see one, two or even three PE plants in the configuration.

Similarly, at least one PP plant always fits in the configuration. The integration of petro-FCC with a PP plant has become the norm in many refineries that do not want to venture into other petrochemicals. Similarly, the combination of propane dehydrogenation (PDH) and PP plants has also risen in last two decades. This has increased the trend of one or two large PP plants with the ever-expanding PP market.

With the inclusion of multiple PE/PP plants to match the olefins balance, the single-line capacities of PE/PP plants have risen to 450,000 tpy–550,000 tpy. Until about 2010, PE/PP plant capacities of 300,000 tpy–400,000 tpy were considered standard for a large plant. However, PE/PP technology licensors have kept up the pace by offering capacities in the range of 450,000 tpy–650,000 tpy, depending on the product mix.

This article will explore the determining factors for different premium technologies that might limit single-line capacities, as well as potential challenges for future capacity increases.

## PE/PP CAPACITY PERSPECTIVES

PE/PP plant capacities must be looked at from two different perspectives: one perspective is the technology licensors', and the other is the viewpoint of owners-producers that do not license technology.

Owners-producers develop, build and operate their own plants, which can be in one or many locations and are either owned outright or part of a JV. These PE/PP producers do not license the technology. Capacities of these plants have mainly been built in a modular way and are not necessarily comparable with other plants; therefore, these plants are outside the purview of this article.

Two types of technology licensors operate: those that develop, build and operate their own plants, either at their own locations or at a JV's locations; and those that, at this stage, do not operate their own plant or do not possess their own plants. Obviously, either they or their predecessors of these technologies have pioneered, developed, built and operated their own plants in the past.

When licensors license the technology know-how to licensees, they design the plants to meet the performance guarantees as per license agreements. Under such situations, the licensors will design the plants with reasonable design margins of typically 10%–15%. Therefore, if a plant is guaranteed for 100% capacity, the licensor will design the plant for 110%–115 % of that capacity.

This article addresses licensed technologies about which some information is available in public domain.

**PE technologies: Present and emerging scenarios.** PE has been the leading thermoplastic resin in the market for the last seven decades. While the low-density polyethylene (LDPE) produced by high-pressure technology emerged in the

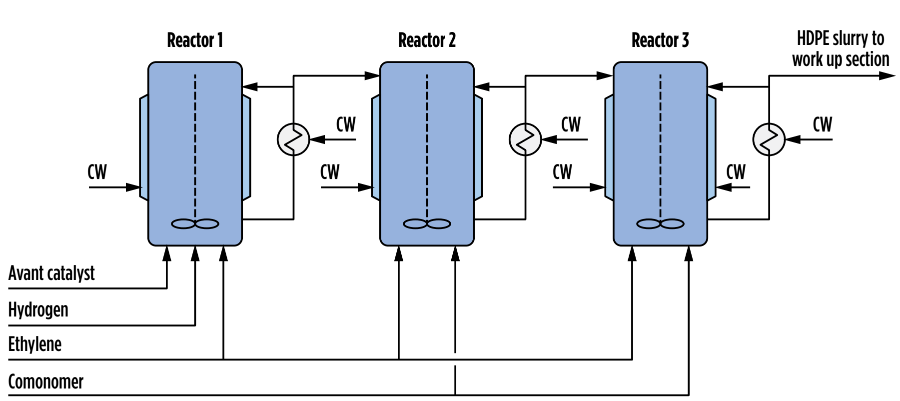


FIG. 1. Polyethylene: CSTR—Hostalen ACP process flow. Source: LyondellBasell.

1930s, the real breakthrough took place in 1950s with the advent of high-density polyethylene (HDPE) produced by low-pressure processes using Ziegler catalysts or chromium-based catalysts.

Since then, evolving HDPE technology has brought radical changes to the way PE has been used for diverse applications. HDPE, with a broad range of applications, is normally produced either by a slurry process or by a fluidized-bed gas phase process. The slurry process is further divided in two processes: continuous flow stirred tank reactor (CSTR) or loop slurry. Main licensors in this category include LyondellBasell (CSTR)[1], shown in **FIG. 1;** Mitsui Chemicals (CSTR)[2], shown in **FIG. 2;** and Chevron Phillips Chemicals (loop slurry)[3], shown in **FIG. 3,** using one or multiple reactors. The single-train capacities with these technologies have settled in the 400,000 tpy–500,000 tpy range to produce especially niche applications in the high molecular weight (HMW) range.

In the 1960s, Union Carbide introduced a fluidized-bed gas phase process (now popularly known as Unipol PE™, shown in **FIG. 4**) to produce HDPE focussing mostly on the lower end of applications in terms of molecular weight. Its successor Univation[4] is presently the only licensor in this technology segment after Ineos discontinued its licensing business. Unipol PE™ offers either Ziegler-Natta (ZN) catalysts, chromium-based catalysts or speciality catalysts to cover the entire range of product applications.

In the 60s and 70s, two processes evolved to pioneer a linear low-density polyethylene (LLDPE) solution and fluidized-bed gas phase process. However, this solution technology that revolutionized LLDPE is practically extinct due to higher capital cost and cumbersome operations compared to competing technologies and does not appear to be prospective for any new plant. No new plant based on this technology has emerged in the last decade.

Univation, successor of Union Carbide, has meanwhile maximized the concept of "swing" operation to produce LLDPE and HDPE in a single reactor. Here again, after the exit of Ineos from the licensing scene, Univation is enjoying a virtual monopoly in this category. The main benefit of Unipol PE™ is its capability to offer large single-line capacities up to 650,000 tpy, and potentially even beyond. The technology offers ZN catalysts, chrome-based catalysts and metallocene catalysts for LLDPE variety, thereby promising everything from LLDPE to high-end HDPE grades all in one reactor.

Main polyethylene technology licensors have been summarized in **TABLE 1.**

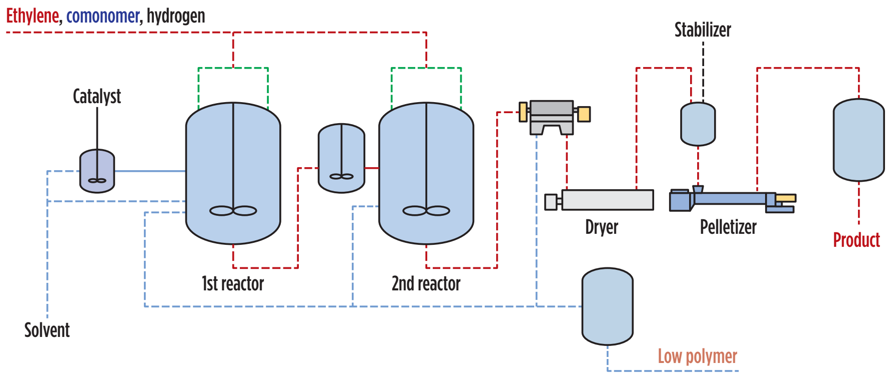


**FIG. 2.** Polyethylene: CSTR—Mitsui CX process flow. Source: Mitsui Chemicals.

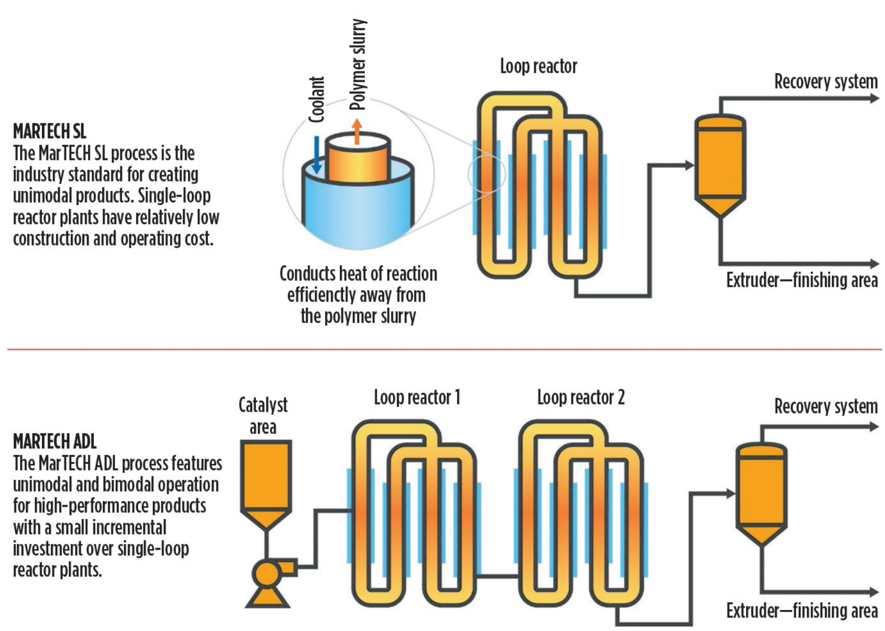


**FIG. 3.** Polyethylene: Loop slurry process. Source: Chevron Phillips Chemicals.

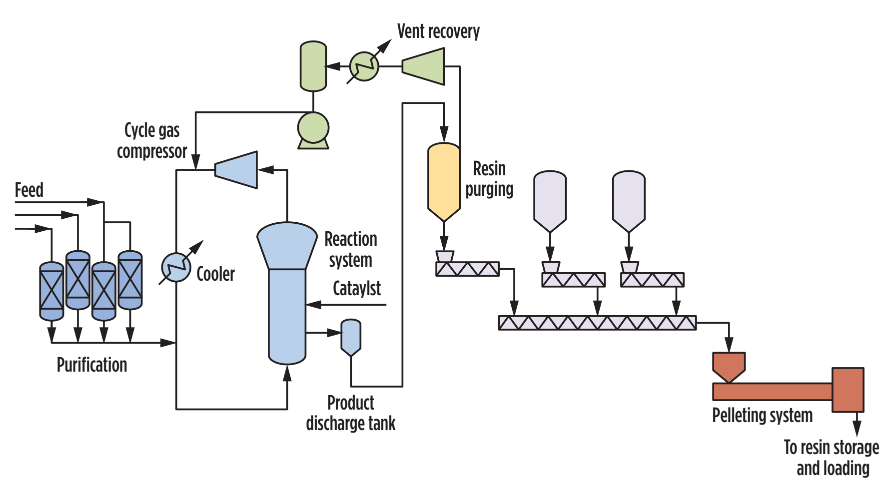


**FIG. 4.** Polyethylene: Univation (UNIPOL) fluidized-bed gas phase.

**PP technologies: Present and emerging scenarios.** PP is the second most popular thermoplastic resin in terms of volume and applications, and is manufactured mainly by loop slurry, fluidized-bed gas phase and stirred-bed gas phase processes.

Loop slurry technology is offered by LyondellBasell (Spheripol™)[5], shown in FIG. 5, and Mitsui Chemicals (Hypol™)[6]; fluidized-bed gas phase technology is offered by W. R. Grace and Co. (Unipol PP™)[7]; and stirred-bed gas phase technology is offered by Lummus Novolen (Vertical Stirred Gas Phase)[8], shown in FIG. 6. Ineos offered a horizontal-stirred gas phase process until it exited the licensing business in 2016–2017. With the exit of Ineos, the competition has been reduced mainly to LyondellBasell, W. R. Grace and Co., Lummus Novolen and Mitsui Chemicals. Except Mitsui Chemicals, all other licensors offer single-line capacity in the range of 450,000 tpy–600,000 tpy, depending on product mix. Mitsui Chemicals is expected to offer competing single-line capacities in the future.

Homopolymer PP, which constitutes the bulk of volume (approximately 70%–75%), is produced in a single-stage reactor or two parallel reactors, depending on the technology. Random copolymers PP, which constitutes about 5%–10% of volume, is also produced in a single-stage reactor or two parallel reactors like homopolymer PP using ethylene as co-monomer.

Impact copolymers using ethylene as co-monomer, constituting 20%–25% volume, are produced in series or cascade mode with the second reactor being a fluidized-bed gas phase reactor, as in the case of LyondellBasell, W. R. Grace and Co. and Mitsui technologies, and a vertical-stirred bed reactor for Lummus Novolen. All of these technologies offer their ZN-based proprietary catalyst systems with either proprietary external donors or commercially available donors. Though homopolymer PP dominates the volume, impact copolymers are making strong inroads to newer and innovative applications. Ter-polymers are also emerging for specialized applications like low-temperature heat sealing.

PP technology licensors have been summarized in TABLE 2.

**LDPE technologies: Present and emerging scenarios.** Demand for LDPE products, once thought to be fading, has re-emerged strongly in the last two decades. LDPE is manufactured either by a high-pressure tubular process or high-pressure autoclave process. Technology licensors for LDPE include LyondellBasell, ExxonMobil and Sabtec. In plants built over the last two decades, LDPE is usually manufactured exclusively by a tubular reactor process. Autoclave technology is used to produce only co-polymers like ethylene vinyl acetate (EVA), which has been growing in importance for speciality applications. Design capacities of tubular or autoclave processes are driven solely by high-pressure engineering capability. The tubular process offers up to 450,000-tpy

**TABLE 1.** Polyethylene technologies

| Licensor | LyondellBasell | Mitsui Chemicals | Chevron Phillips Chemicals | Univation |
|---|---|---|---|---|
| Technology | CSTR | CSTR | Loop slurry | Fluidized-bed gas phase |
| | ACP Hostalen™ Multi-Modal | CX™ Bi-Modal | Martech SL™/ Martech ADL™ Uni-Modal/Bi-Modal | Unipol PE™ Uni-Modal/ Bi-Modal |
| Operates own/ JV plants | Yes | Yes | Yes | No |
| Largest known single-line capacity | 400,000 tpy | 400,000 tpy | 550,000 tpy | 550,000 tpy |

**TABLE 2.** Polypropylene technologies

| Licensor | LyondellBasell | Mitsui Chemicals | W. R. Grace and Co. | Lummus Novolen |
|---|---|---|---|---|
| Technology | Loop slurry | Loop slurry | Fluidized-bed gas phase | Stirred-bed gas phase |
| | Spheripol™/ Spherizone™ | Hypol II™ | Unipol PP™ | Lummus Novolen™ |
| Operates own/ JV plants | Yes | Yes | No | No |
| Largest known single-line capacity | 450,000 tpy | 400,000 tpy | 550,000 tpy | 500,000 tpy |

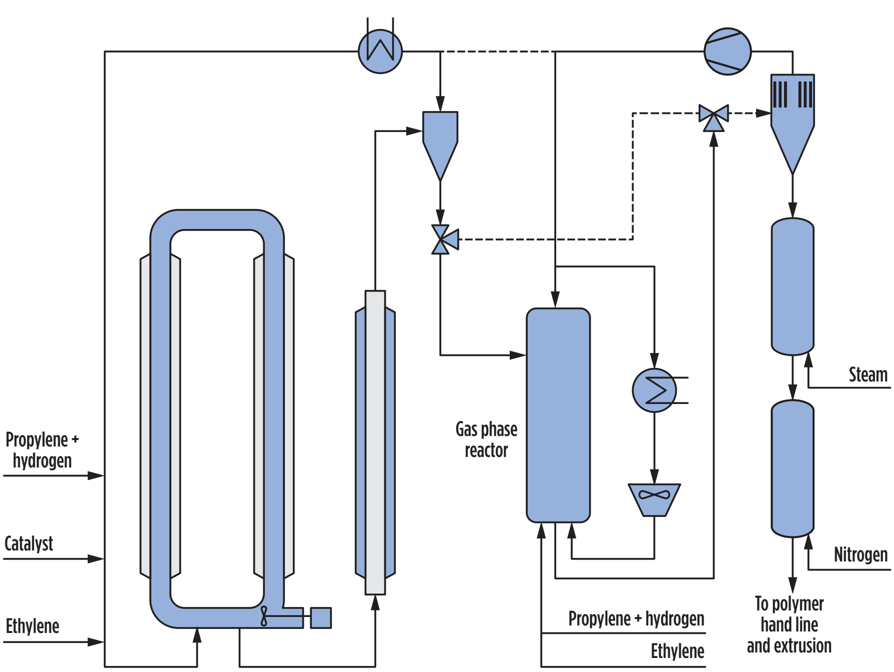


**FIG. 5.** Polypropylene: Spheripol loop slurry simplified process flow. (Source: LyondellBasell)

plants and the autoclave process offers up to 150,000-tpy capacity for EVA. The tubular process can also provide up to 20% EVA copolymers, while the autoclave process can produce up to 40% EVA. No further advancement in single-line capacity, either by tubular or by autoclave processes, is expected in near future.

## DRIVING REACTOR CAPACITIES

Two primary areas of continuous innovation and advancement have driven the capacity of the single-reactor line. One area is specific to process technology and the other is common for all technologies.

**Reactor technology.** The specific area pertinent to each technology or each licensor is reactor technology. Regardless of technology, the common area is extruder or pelleting that converts resin powder to product pellet through melting, which is more rheological scale-up and machinery advancement in the pelleting equipment.

Ethylene and propylene polymerization reactions are highly exothermic. The reaction demands heat removal as fast as possible to control the reactor temperature and avoid "hot spots" or localized reactions. Reactor pressure is usually held constant and temperature is controlled, depending on the type of polymer being produced, within +/–1°C. The heat is removed either by jacket cooling or by external coolers, or a combination of both, and the mechanism varies from technology to technology.

The role of the catalyst system is unique in polymerization in that catalyst enables the reaction and dictates the polymer characteristics or polymer product properties. This mandates certain minimum residence time in the reactor, which again varies for each technology. Therefore, reactor size or volume is primarily determined considering the aspects of heat transfer, mass transfer and reaction kinetics.

Catalyst yield or catalyst productivity—as defined by tons of polymer produced per kg of catalyst (metric t/kg)—has been the single most important factor driving PE/PP technology advancement. Though this is not the theme of this article, it should be noted that catalyst productivity through continuous research and innovation has advanced multi-fold to a sufficiently high level to enable the large single-reactor line capacities that are now prevalent. Key factors contributing to reactor capacity differ for each type of technology.

**CSTR technologies for PE.** Two main factors ae involved: reactor volume and reactor agitator. The reactor is a pressure vessel where pressures are moderate, so design for a given volume is not a controlling factor. It is the agitator system design for large vessels that is critical.

Ethylene polymerization reaction is heterogeneous, involving a combination of gas (ethylene, hydrogen and co-monomer), liquid (co-catalyst and hexane) and solid (catalyst particles and resin). The agitator design must meet the following minimum criteria:

- Fast or almost instantaneous dispersion of gas into the reaction medium
- Fast reaction velocities
- Solid-liquid suspension consistency
- Heat transfer or heat removal as fast as possible for close temperature control.

Intelligent balance of radial and axial agitation pattern is the key to agitation design. The agitator scale-up has become more predictable and accurate with the advent of computational fluid dynamic (CFD) analysis; nevertheless, experience-based empirical scale-up along with CFD analysis dominates the design. The established vendors have lived up to the challenge to scale up to large capacities, including designing robust agitator sealing systems. A minimum of two reactors are in operation either in parallel or cascade mode. Hostalen ACP technology offered by LyondellBasell employs three reactors.

Multiple reactors are intended for manipulation of product properties, but also to assist in reducing the throughput per reactor. Individual jacketed reactors have reached the size of approximately 350 m³–400 m³. Most of the heat is removed by external coolers by circulating slurry pumps.

**Loop slurry reactor technologies for PE.** Again, reactor volume and a loop circulation pump are the two main factors. The reactor is in the form of long vertical vessels called reactor loops or legs. The heterogeneous reaction includes gas (ethylene, co-monomer and chain terminating agent), liquid (iso-butane) and solid (resin and catalyst). The residence time or volume and heat transfer aspects are determined by L/D ratio. This has given rise to multiple legs or loops operated in series or parallel with multiple circulation pumps. Therefore, increasing the diameter and L/D ratio per loop combined with multiple loops has enabled the reactor scale-up to a higher capacity. More than the reactor volume, the scale-up of the slurry circulation pump has been the challenge for speciality pump manufacturers. The circulation pump must have a large flowrate operating with low $\Delta P$—but at high suction pressure—to maintain high velocity to sustain high slurry consistency or high solid-to-liquid ratio. A mechanically challenging task has been to provide a sealing system to handle ethylene and light hydrocarbons like isobutane, in addition to scale-up of pump impeller design.

The established vendors have overcome this problem gradually to supply the

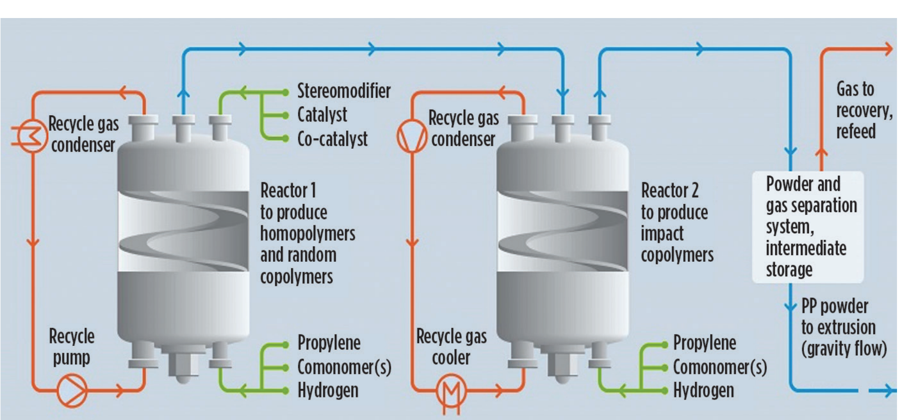


**FIG. 6.** PP: Stirred-bed gas phase process. Source: Lummus Novolen.

required large-capacity pumps that operate at severe conditions.

**Loop slurry reactor technology for PP.** The previous discussion also holds true for PP technology, in principle. The difference lies in the fact that reactant propylene acts as a diluent for the slurry. The fluidized-bed gas phase reactor is used in cascade mode to produce impact copolymers.

**Fluidized-bed gas phase reactor for PE.** Here, the reaction is carried out in a fluidized bed and the resultant heat of the reaction is carried to the external cooler by carrier gas or recycle gas via a recycle gas compressor.

Part of the heat of the reaction is also removed by evaporative cooling using inert iso-pentane in the cycle gas, which is also termed as condensing mode. The reactor has two parts: one is the cylindrical vertical shell that holds the fluidizing resin bed, and the second consists of a large dome-shaped vessel on top of the cylindrical shell, which disengages the cycle gas from the resin powder before leaving the reactor.

The volume of the dome is very large compared to the shell to enable separation of cycle gas from any resin particles. The fluidizing bed volume ensures the minimum residence time and acts as a heat dissipating medium. The volume of the cylindrical shell determines the minimum residence time. The fluidizing velocity must be high so that heat dissipation is fast for accurate and close temperature control, and at the same time allowing good mixing of catalyst and reactants in the resin bed so that the fluidizing bed nearly approximates the CSTR, although it is not comparable to a CSTR.

Therefore, reactor volumes guided by L/D ratio and fluidizing velocity, as well as the corresponding volume of the dome, have grown multi-fold over the years to match the single reactor capacities. The single reactor is used for producing bimodal HDPE products using radical innovations in catalyst technology.

Recycle gas compressor capacities have kept pace with the required gas flowrates to match reactor capacities. The established vendors have scaled up compressor performance, which requires very high flowrates at low $\Delta P$ but at a high suction pressure and an associated robust mechanical sealing system to handle ethylene, co-monomers, hydrogen, etc.

**Fluidized-bed gas phase reactor for PP.** The above discussion also holds true for PP technology, in principle, but with some variance. Excess reactant propylene is used for evaporative cooling—termed as condensing mode operation—to enhance the heat removal capacity in the given reactor. While some differences in mechanism of ethylene and propylene fluidization exist, the overall design principles are mostly the same.

**Vertical-stirred bed gas phase reactor.** Here, the reactor is a CSTR type without any diluent, with the resin bed being kept in motion by a low-carrier gas velocity.

**TABLE 3.** Main scale-up factors for PE/PP technologies

| | Fluidized-bed gas phase | Stirred-bed gas phase | CSTR | Loop slurry |
|---|---|---|---|---|
| **Polymerization** | PE/PP | PP | PE | PE/PP |
| **Medium** | Resin bed | Resin bed | Slurry | Slurry |
| **Diluent** | Not applicable | Not applicable | Hexane | Liquid propylene for PP |
| | | | | Isobutane for PE |
| **Carrier** | Cycle Gas | Cycle gas | Slurry | Slurry |
| **Reactor vessel** | Vertical cylindrical + dome | Vertical cylindrical | Vertical cylindrical | Multiple long vertical cylindrical called loop or leg |
| **Agitator** | Not applicable | Yes | Yes | Not applicable |
| **Cooling jacket** | No | No | Yes | Yes |
| **External cooler** | Yes | Yes | Yes | No |
| **Recycle gas compressor** | Yes | Yes | No | No |
| **Scale-up factors** | | | | |
| **Residence time** | Yes | Yes | Yes | Yes |
| **Reactor volume** | Yes | Yes | Yes | Yes |
| **Reactor diameter** | Yes | Yes | Yes | Yes |
| **L/D ratio** | Yes | Yes | Yes | Yes |
| **Space time yield** | Yes | Yes | Yes | Yes |
| **Velocity** | Yes: Well above fluidization velocity | Not a factor | Yes: Slurry circulation pump | Yes: Slurry circulation pump |
| **Specific agitator power** | N/A | Yes | Yes | N/A |
| **Circulation pump** | N/A | Yes | Yes | Yes |
| **Recycle gas compressor** | Yes | Yes | N/A | N/A |
| **Jacket heat transfer area** | N/A | N/A | Yes | Yes |
| **External cooler heat transfer area** | Yes | Yes | Yes | N/A |

The excess reactant propylene is also used for evaporative cooling, thereby serving as evaporating liquid, as well as carrier gas.

Two primary areas have driven the capacity of single-reactor lines: one is specific to process technology and the other is common for all technologies.

The high velocity is not required because agitation has substituted the task of fluidization. The mechanical mixing of catalyst in the resin bed improves the mixing performance, as well as reaction and heat dissipation. The carrier gas flowrates are quite small because there is no fluidization. The minimum residence time is handled by reactor volume guided by L/D ratio.

Again, as with other CSTR technologies, the design of the pressure vessel has not been much of a problem, unlike the scale-up of the agitator design. However, unlike CSTR, this agitator does not require robust mixing by radial and axial movements, but rather a gentle agitation intensity. The agitator design is different here vs. CSTR in the sense that here, primarily gas-solid mixing is involved with a very small liquid phase.

Gas-solid mixing has been more empirical than the CSTR agitator. Again, the advent of CFD analysis has helped perfect the proprietary agitator design. Lummus Novolen has entered the licensing business relatively late but has quickly accelerated the capacity scale-up to the global standard with the assistance of established agitator vendors.

The essence of reactor scale-up challenges are summarized in **TABLE 3.**

**Extruder or pelleting package.** An extruder is an essential unit in any PE and PP plant, irrespective of technology or licensor. The polymerization reaction system is followed by resin degassing, resin powder treatment, polymer additive mixing and, finally, an extruder or pelleting unit. The extruder is either resin powder fed or resin melt fed. Except LDPE and solution PE technology, all other technologies have resin powder fed extruders. Similarly, all these extruders are twin-screw extruders. An extruder unit is a package item that consists of several components, including:

- Main motor and main gearbox
- Processing section or main extruder consisting of a combination of several proprietary screw elements, gear pump, die plate, underwater pelletizer, pellet dryer, etc.

When scale-up of extruder capacities are discussed, all associated components are automatically included. Each of the components must match the overall extruder capacity. In fact, when the main processing section is specified, the capacity-controlling downstream units (e.g., gear pump, die plate, pelletizer, pellet dryer) must be 10%–15% higher capacity than the main processing section.

Using the same analogy, the extruder package must be 10%–20% higher capacity than the reaction section. This ensures that a minor maintenance shutdown in the extruder area of 4 hr–8 hr does not affect reactor operation; when the extruder is restarted, it can operate at a higher capacity than the reactor to bring down upstream resin inventory to normal levels and set the plant again on steady-state operation. As a guiding principle of plant design, any equipment downstream of the reaction area is not expected to become capacity controlling or limiting.

**TABLE 4** illustrates how the extruder packages and pellet conveying (or pneumatic conveying packages) are designed based on plant capacity or reactor capacity. It should be noted that extruder and conveying packages are typically designed for either reactor:extruder:conveying = a 100:120:140 or 100:110:120 capacity ratio with respect to the reactor.

100:120:140 is a more conservative approach and is usually desired and practiced to its fullest extent for any plant. However, the 100:110:120 pattern is sometimes selectively chosen to optimize the economics of the plant or keep the initial CAPEX to a lower level for a higher capacity of > 500,000 tpy, considering product mix.

**Extruder capacity.** The extruder package is the single costliest item in the whole PE/PP cost estimate. An extruder package occupies a large footprint and adds complexity to the plant design in addition to higher plant CAPEX. The trend is to install only one extruder of equivalent capacity rather than dividing it into two smaller extruder packages.

An exception is made only in those PE plants where a high proportion of

**TABLE 4.** PE/PP plants: Extruder and pneumatic conveying capacity vs. reactor capacity

| Plant capacity | 1,000 tpy | | 400 | 500 | 600 | 700 | 800 |
|---|---|---|---|---|---|---|---|
| **Conservative case** | | | | | | | |
| Reaction area | tph | 100% | 50 | 62.5 | 75 | 87.5 | 100 |
| Extruder package | tph | 120% | 60 | 75 | 90 | 105 | 120 |
| Pneumatic conveying | tph | 140% | 70 | 87.5 | 105 | 122.5 | 140 |
| Pellet blending and storage silos | tpd | | 1,200 | 1,500 | 1,800 | 2,100 | 2,400 |
| **Minimum case** | | | | | | | |
| Reaction area | tph | 100% | 50 | 62.5 | 75 | 87.5 | 100 |
| Extruder package | tph | 110% | 55 | 68.75 | 82.5 | 96.25 | 110 |
| Pneumatic conveying | tph | 120% | 60 | 75 | 90 | 105 | 120 |
| Pellet blending and storage silos | tpd | | 1,200 | 1,500 | 1,800 | 2,100 | 2,400 |

pipe grades are to be made (e.g., PE 100 pipes or black pipes), thus justifying the installation of a dedicated extruder for pipe grades.

An extruder with a motor or drive of 15 MW–20 MW is considered very large and are typically single-speed extruders because variable-speed extruders with such large motors are extremely expensive. Additionally, the specific energy index of an extruder package defined as Kwh/kg (kilowatt hour/kilogram) of capacity becomes high and cost prohibitive due to energy or electricity costs.

The extruder capacity is nominally specified by screw diameter. Established extruder package vendors have successfully demonstrated operation with 380-mm–450-mm screw diameters with improved filling efficiency. Mega compounders, as they are called, have nearly reached a torque limit for standardized throughput rate with respect to specific energy, especially for LLDPE and HDPE for certain applications at approximately 100 t/hr.

A single extruder capacity up to 100 t/hr is considered a proven and reliable scale-up. This would imply that components downstream of the processing section are designed for a capacity higher than 100 t/hr. Therefore, it is logical that a reactor or plant capacity of 80 t/hr–85 t/hr (or the equivalent to 700,000 tpy) is achievable with operational reliability if an extruder capacity of 100 t/hr is considered a proven scale-up.

**Silos and pneumatic conveying.** Silos are generally made either of stainless steel or Al/Al alloys. These pressure-less vessels are fabricated onsite. The number and size of silos are decided by the client based on their operating philosophy. However, each silo with a capacity of 700 t–1,000 t is common in the industry with 24 hr–48 hr of combined installed storage space for blending and storage determining the number of silos.

**Pellet pneumatic conveying.** Product pellets are pneumatically conveyed from the extruder to blending/storage silos and conveyed from silos to warehouse for bagging or packaging.

The pneumatic conveying of pellets is largely dictated by the principle of velocity to prevent solids from settling during transportation by conveying. Consequently, pipeline sizes increase proportionately to capacity and tend to become large. This is not as evident in straight pipelines; it is piping elements like bends, diverter valves, rotary feeders, etc., that really set the limitation on the confidence of the scale-up and the proven-ness of the components. Generally, pneumatic conveying capacity up to 120 t/hr is considered proven due to reliable scale-up, although capacities up to 140 t/hr appear promising in the future.

**Single-line capacity challenges ahead.** This discussion suggests that single-line capacities for PE and PP plants are expected to range from 500,000 tpy–700,000 tpy at the higher end of the spectrum. Plant capacities will be dictated by product mix and the technology selected. Technology selection is guided by factors that include:

- Flexibility
- Versatility
- Product range
- Catalyst range and productivity
- Product transition time
- Proven capacity of critical equipment, such as the reactor, extruder, etc.

This is of course over and above the usual CAPEX and OPEX parameters.

**Takeaway.** PE faces a greater challenge due to the nature of a wide spectrum of product application, ranging from LLDPE, MLLDPE, MDPE, MM-HDPE and HM-HDPE to bi-modal HDPE. No single technology is expected to encompass the entire range in equal measure for various reasons. This has necessitated licensees to seek a judicious combination of technologies to cover the entire product range, or to focus on a selective product range. The conceivable combinations can be but are not limited to:

- Fluidized-bed gas phase + loop slurry
- Fluidized-bed gas phase + CSTR
- Two fluidized-bed gas phases with swing capability
- Two-loop slurry: SL + ADL.

The single-line capacity of these technologies can be expected to advance to the range of 550,000 tpy–750,000 tpy soon with an emphasis on energy efficient design.

PP has an advantage over PE in that all PP technologies offer a full product range. If ethylene is unavailable, only a homopolymer PP plant can be installed with the provision to add an impact copolymer PP reactor later—Lummus Novolen in an exception, as its two-reactor technology can produce the entire product range. All PP technologies are expected to advance toward a 550,000 tpy–750,000 tpy range soon with an emphasis on energy efficient design.

It is conceivable that the challenge of large-capacity plants can be partly overcome by installing one or more small-capacity, high-value speciality plants that are ethylene or propylene derivatives other than the usual PE and PP. The new trend of diverting part of the ethylene and propylene to produce more value-added speciality chemicals looms bright on the horizon and will emerge sooner than later. Plant size will be conveniently smaller and, accordingly, CAPEX will decrease with relatively higher returns. Elements of business risks accompanying large single-line capacity will be replaced by opportunities. HP


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# Best practices for pygas-based styrene extraction

Pyrolysis gasoline (pygas) is a by-produced fraction of hydrocarbons generated from a steam cracker. Rich in aromatics content, particularly for benzene and toluene, pygas was historically blended into the gasoline pool as a high-octane number component. In this modern era, due to stringent benzene specifications in the gasoline pool and growing demand for the aromatic molecules from petrochemicals and chemicals manufacturing, the majority of pygas is fed to aromatics extraction units for benzene, toluene and mixed xylenes (BTX) recovery.

How much pygas can be by-produced from a steam cracker primarily depends on the type of feed to the cracker and the cracking severity. **TABLE 1** lists the yield of pygas, which is typically defined as the hydrocarbon fraction from $C_5$s to 204°C (399.2°F) end point, from the various cracker feedstocks under relatively high cracking severity.

The composition of pygas consists of more than 200 species, including paraffins, naphthenes, olefins, diolefins, cyclo-olefins, acetylenes, aromatics, alkenyl aromatics and multi-ring aromatics, along with impurities, such as sulfur, nitrogen and chloride. **TABLE 2** indicates a typical hydrocarbon PONA breakdown of pygas when cracking naphtha feed.

To prepare the feed for aromatics extraction, mono-olefins and diolefins in the raw pygas need to be saturated, and sulfur, nitrogen and chloride need to be removed. The practical approach is to hydrotreat the raw pygas in two stages. First-stage hydrotreating saturates diolefins to mono-olefins, and second-stage hydrotreating saturates the mono-olefins and removes sulfur, nitrogen, chloride and other impurities. While the two-stage hydrotreating scheme is widely accepted in the industry, the fractionation scheme for the hydrotreated pygas varies from case-to-case. Fractionation is optimized based on the evaluation of process technology and economic benefits of separating various products. Typically, for the mega-sized liquid-fed steam crackers, the $C_5$ cut and $C_9$+ cut are segregated from the $C_6$–$C_8$ heart-cut since both the $C_5$ and $C_9$+ cuts contain many valuable unsaturated monomers that can be utilized for manufacturing of high-value derivative products. The $C_6$–$C_8$ cut is hydrotreated in two stages and sent for the extraction of benzene, toluene and xylenes. **FIG. 1** presents the integrated process configuration of pygas pre-fractionation, heart-cut two-stage hydrotreating, aromatics extractive distillation and aromatics post-fractionation.

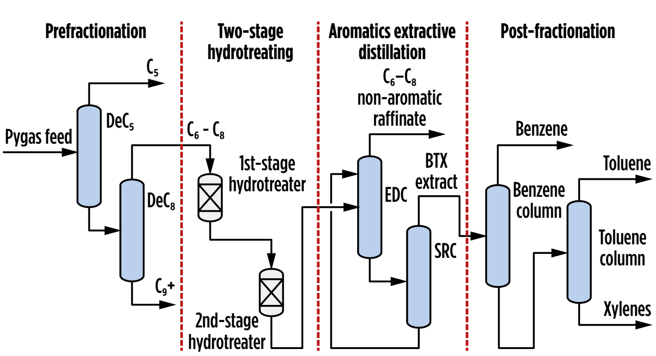


**FIG. 1.** Typical pygas processing scheme.

Steam cracking of liquids such as naphtha, diesel and gasoils co-produces 25 kg–35 kg of styrene monomer (SM) for every ton of ethylene production. This corresponds to about 5 wt% SM in the raw pygas and about 40 wt% SM in a raw $C_8$-rich cut when such a stream is separated. **TABLE 3** provides pygas $C_8$ cut compositions. For a steam cracker producing 1 MMtpy of ethylene, there is the potential to recover up to 35,000 tpy of SM. Historically, styrene recovery from pygas was not considered because the right technology was not available and ethylene crackers were not large enough for recovering styrene at an eco-

**TABLE 1.** Cracker yields

| Cracker feedstock | Amount of feedstock consumed when 1 t of ethylene is produced, t | Amount of pygas by-produced when 1 t of ethylene is produced, t |
|---|---|---|
| Ethane | 1.246 | 0.25 |
| Butanes | 2.47 | 0.162 |
| Light-naphtha | 3.18 | 0.48 |
| Full-range naphtha | 3.423 | 0.764 |
| Atmospheric gasoil | 3.938 | 0.716 |
| Vacuum gasoil | 4.282 | 0.687 |

**TABLE 2.** Typical pygas composition

| Wt% | Saturates | Olefins | Diolefins | Aromatics | Total |
|---|---|---|---|---|---|
| $C_4$ | 0.1 | 0.2 | 0.2 | – | 0.5 |
| $C_5$ | 0.3 | 3.4 | 11.8 | – | 15.5 |
| $C_6$ | 0.7 | 0.9 | 4 | 31.2 | 36.8 |
| $C_7$ | 0.6 | 0.7 | 2.5 | 16.3 | 20.1 |
| $C_8$ | 0.3 | 0.4 | 1.5 | 9.6 | 11.9 |
| $C_9$+ | 1.2 | 0.3 | 3.1 | 10.8 | 15.3 |
| **Total, wt%** | **3.3** | **5.9** | **23** | **67.9** | **100** |

nomic scale. Starting around 10 yr ago, the capacity of an ethylene cracker has become larger—especially those integrated with the refinery—as many streams from refineries (such as LPGs, naphtha and gasoil) could be optimized as feed to the cracker and push the olefins production capacity over 1 MMtpy.

The authors' company began research and development work on recovering styrene from pygas 20 yr ago by combining the knowledge and experience of working with both separation equipment and extractive solvents. An extractive distillation-based process for recovering styrene from pygas was invented and commercialized. This proprietary styrene technology[a] was first licensed in 2005, with the first commercial unit commissioned in 2009. FIG. 2 provides the process flow of the proprietary styrene technology.

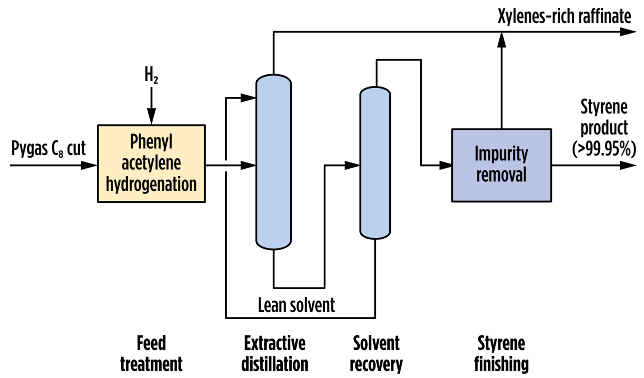


FIG. 2. Process flow diagram of the proprietary styrene technology[a].

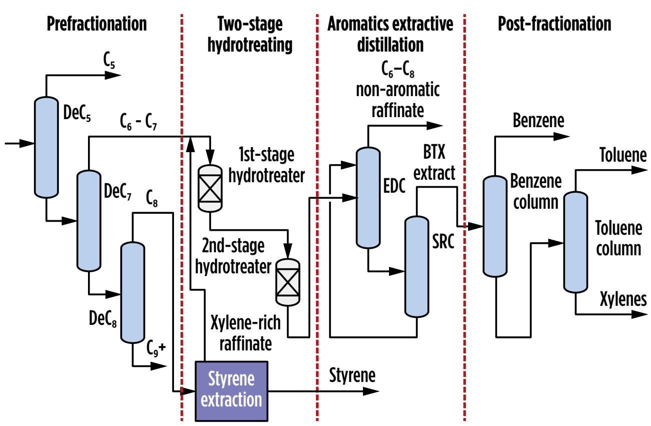


FIG. 3. Pygas processing scheme with styrene extraction.

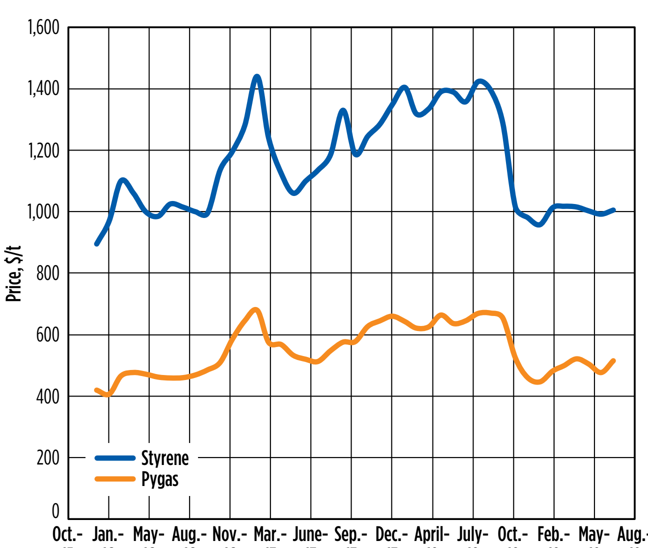


FIG. 4. Price difference between pygas and SM.

TABLE 4 lists the ASTM specification D2827-19 for SM specification. The purity, aldehydes, peroxides, polymer, inhibitor, ethylbenzene (EB), water and color will be tested to meet the requirements specified. The quality of the extracted styrene product complies with the international standard.

The styrene extraction unit is installed before the first-stage pygas hydrotreatment unit in the pygas processing complex. A new deoctanizer column is required to separate a $C_8$-rich cut from the $C_9$+ stream. FIG. 3 shows the integrated pygas processing scheme with both aromatics extraction and styrene extraction. Xylene-rich raffinate from styrene extraction, combined with the $C_6$-$C_7$ cut, is routed to the inlet of the first-stage hydrotreating, then to aromatics extraction for xylenes recovery.

When analyzing the benefits of incorporating styrene extraction into the pygas processing complex, the price spread between raw pygas and SM is one of the most critical considerations. Typically, pygas is valued for its benzene content, with the remainder as a gasoline component. FIG. 4 shows the historical data of price differences between pygas and SM.

Other than the economic benefit of upgrading styrene mol-

**TABLE 3.** Pygas $C_8$ cut composition

| Component | Wt% |
|---|---|
| $C_7$ non-aromatics | 0.23 |
| $C_8$ non-aromatics | 7.87 |
| Toluene | 0.12 |
| P-/m-/o- xylenes | 40.81 |
| Ethylbenzene | 9.95 |
| Styrene | 39.35 |
| Phenylacetylene | 1.2 |
| $C_9$ | 0.46 |
| **Total** | **100** |

**TABLE 4.** ASTM specifications for SM

| Property | Specification | ASTM test method |
|---|---|---|
| Color, platinum/cobalt scale | 15 max | D5386 |
| Styrene purity, wt% | 99.8 min | D5135 or D7504 |
| Aldehydes (as benzaldehyde), wt% | 0.01 max | D2119 or D7704 |
| Peroxides (as $H_2O_2$), mg/kg | 50 max | D2340 |
| Polymer, mg/kg | 10 max | D2121, test method A |
| Inhibitor, mg/kg | 10–15 | D4590 |
| EB, mg/kg | 500 max | D2135 or D7504 |
| Benzene, mg/kg | 1 max | D6229 |
| Appearance | Clear liquid free of sediment and haze at 18°C–25°C (65°F–78°F) | |

ecules from pygas value to monomer value, there are other benefits connected to extracting styrene out of pygas. Styrene saturation to EB contributes a large portion of the exothermic heat in the first stage hydrotreating. With styrene removed from the feed to the hydrotreater, the heat released in the reactor decreases, the tendency of polymerization for styrene and other diolefins lessens, fouling problems become less severe and the pressure across the first-stage hydrotreater drops. With less heat released, the outlet temperature decreases, which means fewer hydrocarbons vaporize, and hydrogen partial pressure in the reactor outlet increases. To maintain the same partial hydrogen pressure in the vapor phase at the reactor outlet, the hydrogen feed flowrate can be reduced to save hydrogen consumption. With a lower vaporization ratio inside the catalyst bed, better distribution and diffusion of hydrogen inside the liquid hydrocarbon can be achieved. The selectivity of diolefins and acetylenes to mono-olefins is improved. Both catalyst regeneration cycle length and service life will be prolonged. All these advantages result in improved operation of the trickle bed reactor. **TABLE 5** summarizes the impact on pygas hydrotreating with and without styrene extraction.

After applying styrene extraction to pygas processing, the pygas-based mixed xylenes product has a different EB content. In general, pygas-based mixed xylenes are not considered as a preferred feed for PX production, as the EB content is much higher than reformate-based mixed xylenes. This relegates the pygas-based mixed xylenes to solvent use or gasoline blending. After incorporating styrene extraction within the pygas processing scheme, the EB content in pygas-based mixed xylenes drops significantly as a result of avoiding the styrene conversion into EB in the hydrotreater. Typically, the EB content in pygas-based mixed xylenes will be in the range of 20%–30%, with styrene extraction compared with 50%–60% when styrene is not extracted. EB is a molecule that imposes challenges for PX production. Although EB can be reformed to PX through isomerization, the capital expenditure and operating expenditure associated are significantly higher. Any decrease of EB content in the feed to the PX complex means a higher efficiency in terms of per ton of PX manufactured. This is the reason that pygas-based, EB-rich mixed xylenes are not a preferred feedstock for a PX complex.

The integrated refining/petrochemical complex consistently shows higher profitability than the corresponding standalone refinery and ethylene cracker. The refinery sends gasoil, naphtha and LPG to the steam cracker as the feedstock for light olefins production, and receives pyrolysis fuel oil, pygas, $C_4$ raffinate and hydrogen for transportation fuel production. Pygas-derived styrene and the resultant xylenes add another level to improve the integration of processing facilities. Under this scenario, pygas-derived BTX will be supplemental to reformer-derived BTX feeding the aromatics complex for PX production. With the proprietary styrene technology, an integrated refinery/petrochemical complex can produce SM in addition to p-xylene and benzene as aromatic products.

At present, seven proprietary styrene processing units have been licensed, with the largest having a capacity of 80,000 tpy (**FIG. 5**). HP

**TABLE 5.** Summary of the impact on pygas hydrotreating with and without styrene extraction

| | Before proprietary styrene process[a] | After proprietary styrene process[a] |
|---|---|---|
| RIT/ROT for first-stage reactor, °C | 55/102 | 62/85 |
| RIP/ROP for first-stage reactor, kg/cm²g | 27/26.2 | 27/26.5 |
| Recycle/fresh ratio for first-stage reactor | 2.8 | 2.3 |
| $H_2$ consumption over feed, kg/t | 10 | 6.5 |
| Catalyst regeneration cycle, yr | 2 | 4 |
| Catalyst life, yr | 4 | 7 |

**FIG. 5.** Commercial styrene unit—utilizing the proprietary styrene process—operating in China.

**NOTES**

[a] Sulzer GTC Technology's GT-Styrene[sm] process

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# A novel polymerization retarder with boosted performance and improved handling characteristics

Styrene monomer production plants continually face a need to slow polymerization rates in the purification section of the production process to increase monomer production rates, reduce polymerization/fouling and tar formation, and protect against equipment plugging, especially in the event of unplanned shutdowns.[1,2] This protection is typically accomplished by the application of polymerization retarders and true polymerization inhibitors, two related chemistry classes that slow the propagation of polymers.[1,2,3]

The polymerization of styrene monomer begins with the formation of styryl radicals. Under anaerobic conditions, styryl radicals can go on to react with molecules of styrene, forming larger radicals that can then further react with styrene molecules, and the chain can continue until the radicals react with another radical, terminating the reaction.[4,5,6,7] Under aerobic conditions, oxygen will react with styryl radicals at a much faster rate than styryl radicals can react with styrene molecules, thus forming styryl-peroxy radicals. When styryl-peroxy radicals are present, many more propagation reaction paths exist.[8,9,10,11]

A true polymerization inhibitor is defined as a chemistry that completely halts the rate of polymerization while it is present in the monomer solution. The inhibitor is consumed by the process of inhibition, and when fully consumed, the rate of polymerization proceeds at an uninhibited rate. A polymerization retarder, by contrast, does not halt polymerization, but rather it only slows the polymerization rate. Generally, the retarder is not consumed during the retardation process, or is regenerated so that it can continue to react.[3]

The gold standard retarders in the styrene monomer industry for the many years have been a series of "dinitro" compounds, such as 2,6-dinitro-p-cresol (DNPC; CAS Registry Number 609-93-8); similar compounds in this family that have been used as retarders throughout the years include 2,4-dinitrophenol (DNP; CAS Registry Number 51-28-5), 2,4-dinitro-o-cresol (DNOC; CAS Registry Number 534-52-1), and 2,4-dinitro-6-sec-butylphenol (DNBP; CAS Registry Number 88-85-7).[12] While the cost performance of these retarders is well established, so is the toxicity and potential for environmental, health and safety related events, including personnel exposures and environmental releases.[13,14] Even if no unexpected events take place, during turnaround and maintenance operations, costly decontamination procedures are required before equipment that has been exposed to one of these dinitro compounds can be opened for inspection, cleaning or repair.

As an alternative to these toxic dinitro compounds, several process additive companies have developed a first generation of environmentally friendly retarders, most of which are based on quinone methide type chemistries.[15] When these retarders were initially introduced to the market more than 20 yr ago, they showed equivalent performance to the incumbent dinitro compounds at lower dosages; however, the cost was often two to three times higher than the cost of the incumbent dinitro compounds, frequently making these environmentally friendly solutions uneconomical. With time, however, production processes were optimized and raw material costs were controlled, allowing these retarders to become cost competitive. Advantages, such as lower personnel exposure risks, reduced wastewater treatment system impacts, reduced decontamination costs before turnarounds, and reduction in nitrogen oxide ($NO_x$) generation potential from tar burning, can now be realized through this first-generation, environmentally friendly retarder.

This type of first-generation, environmentally friendly styrene retarder has proven success. One service provider[a] that has reported operating nine applications globally has been feeding this type of process additive since the early 2000s. A typical example of a styrene plant's experiences with this type of retarder is described here.

**Case study.** A world-scale styrene plant of a major design[b] was feeding DNBP along with a true inhibitor. Due to a desire to transition away from this toxic retarder, the facility trialed a first-generation, environmentally friendly retarder. The trial was designed to systematically switch the unit from a treatment program consisting of DNBP with a true inhibitor to a treatment program employing a first-generation, environmentally friendly retarder in combination with two different true inhibitors in three phases.

The first phase (denoted as "Phase 1" in **FIGS. 1–3**) introduced a new true inhibitor (Inhibitor 1) at a slightly lower dosage to the previously used inhibitor; the second phase (denoted as "Phase 2" in **FIGS. 1–3**) introduced a second true inhibitor (Inhibitor 2) to optimize inhibitor cost performance. During this second phase, the combined dosage of the true inhibitors was lower than the dosage of the pre-trial inhibitor dosage. **Note:** Due to this plant's unique recycle stream from a downstream polystyrene unit, a spike in the polymer numbers was seen during this phase, caused by an abnormal polymer recycle event.

The third phase (denoted as "Phase 3" in **FIGS. 1–3**) involved

switching the unit from DNBP to the first-generation, environmentally friendly retarder. During this phase, the plant was able to further reduce dosages of the retarder and inhibitor components so that the combined program was the most cost-effective

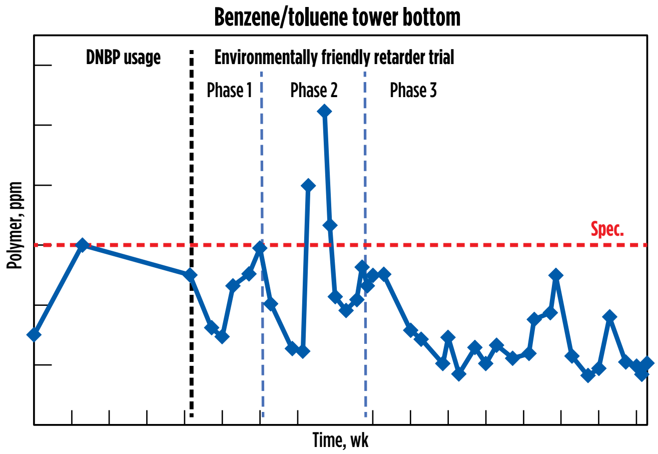


**FIG. 1.** The polymer level trends vs. time in the benzene/toluene tower bottom during various stages of the trial.

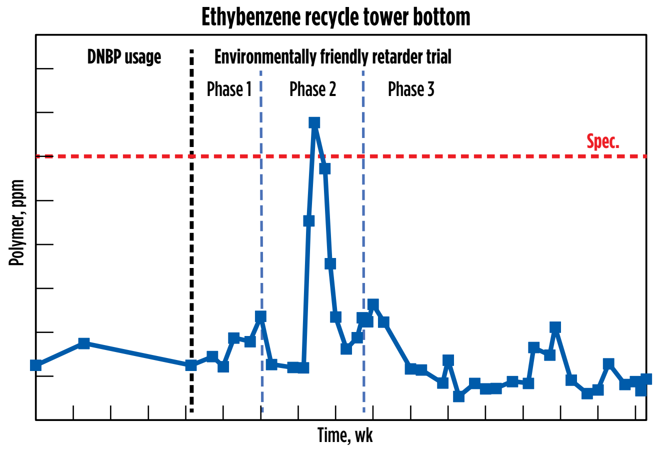


**FIG. 2.** The polymer level trends vs. time in the ethylbenzene recycle tower bottom during various stages of the trial.

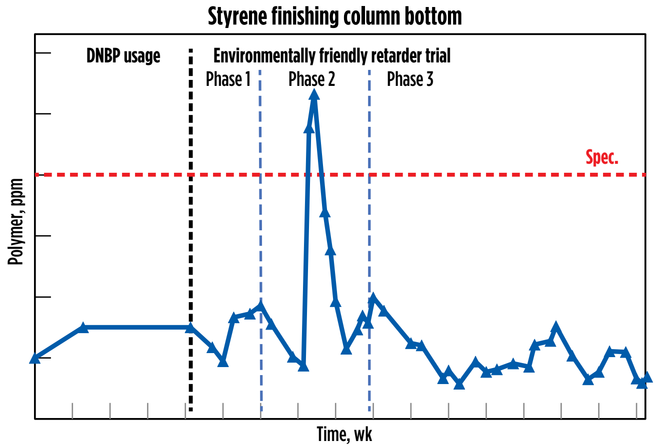


**FIG. 3.** The polymer level trends vs. time in the styrene finishing column bottom during various stages of the trial.

solution, while also allowing the plant to achieve the key polymer level requirements in its benzene/toluene, ethylbenzene recycle and styrene finishing column bottoms, and reduce the overall toxicity profile of its treatment program. **FIGS. 1–3** illustrate the polymer levels in each of these three towers, with each phase of the trial denoted.

Now, a second-generation, environmentally friendly retarder[16], an enhanced quinone methide chemistry, has been developed that, based on laboratory testing, further increases the cost effectiveness of this type of retarder treatment compared to the dinitro incumbents. In addition, it offers improved low-temperature handling compared to the first-generation retarder. Through a patented[16] synergistic relationship between a hindered phenol and a quinone methide type chemistry, the efficacy of the retarder is boosted to a significant level. Beyond the increased cost effectiveness and improved handling, this second-generation product adds the presence of an antioxidant type chemistry to the list of benefits, allowing for increased efficacy even if oxygen is present in the system. The booster component is nitrogen free, allowing the overall nitrogen content of the retarder program to be further reduced compared to the first-generation, environmentally friendly retarders, which were already low in nitrogen content.

**Laboratory evaluation.** The laboratory evaluation of these materials involved an anaerobic reflux of various aliquots of reagent-grade styrene monomer mixed with 100 ppmv of reagent grade divinylbenzene (DVB; CAS Registry Number 1321-74-0), the stock tert-butyl catechol inhibitor having been removed by filtering the combined solution through a column of basic alumina. Before beginning the reflux, each aliquot of styrene monomer was placed in a test tube and dosed with one of the various treatments; the tubes were then sealed with rubber septa.

Each tube was then purged of oxygen by using a pair of non-coring needles to pierce the septa—one needle to vent the tube, and the other used to bubble argon through the tube for a period of 3 min. After sparging, the pair of needles were removed, and the test tubes were placed in a preheated, temperature-controlled, hot oil bath. At periodic intervals, 2.5-ml aliquots of the solutions in each tube were sampled through the septa via the use of a non-coring needle/syringe. The 2.5-ml sub-samples were then injected into 47.5 ml of reagent grade methanol to precipitate any polymer that had been formed. The methanol and styrene mixtures were allowed to sit for at least 16 hr (allowing for full precipitation) before being filtered through pre-weighed glass fiber filters, which captured the polymer formed in the solutions. The filters were dried in a drying oven and then weighed.

**Results and discussion.** This patented technology, which makes use of a synergistic relationship between quinone methides and hindered phenols, is, on an equivalent cost basis, 44.6% more effective than quinone methide alone over a period of 90 min and 55.5% more effective than quinone methide alone over a period of 240 min. When compared to DNBP on a similar active basis, this new combination of quinone methide and hindered phenol is approximately 20% more effective than DNBP over a 90-min period (4% more active material, but a 21.9% improvement in performance). A chart showing the percent polymer formed overtime in styrene monomer treated with the various retarder treatments discussed is shown in **FIG. 4.** HP

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[b] Badger Licensing LLC


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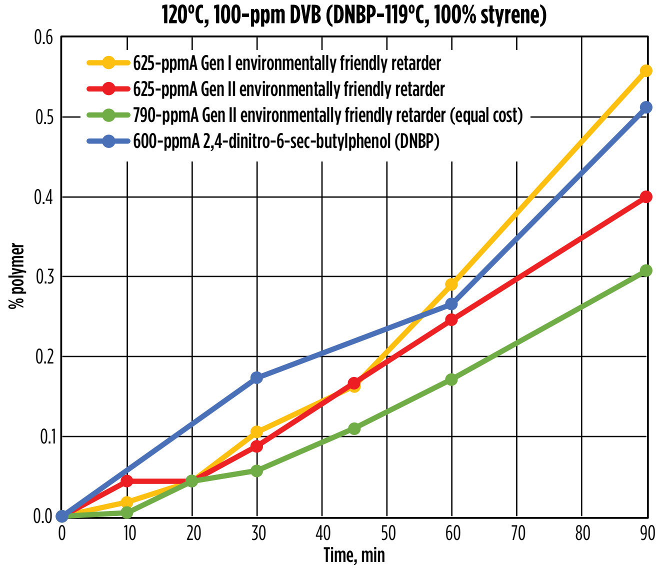


**FIG. 4.** The amount of polymer formed over a period of 90 min under the following conditions: 120°C; pure styrene monomer + 100-ppm divinylbenzene; TBC removed.


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# Specifying internals in sour water strippers—Part 2

Sour water stripping is a common process in petroleum refineries and other processes where hydrogen sulfide ($H_2S$) is present. While not a revenue generator, the sour water treating system is a critical unit operation and can be a significant bottleneck to facility production rates—especially if it is not adequately sized, or if it is forced to operate at partial loads due to maintenance issues. As a result, a balance must be struck between minimizing capital costs, while still providing a reliable and flexible sour water treating system.

Part 1 of this series, published in the February issue of *Hydrocarbon Processing*, provided a general introduction to sour water strippers (SWSs), a process flow diagram for a SWS, an overview of the auxiliary separation equipment needed to remove hydrocarbons and other contaminants from the sour water prior to the stripper, and also reviewed the design of SWS columns containing trays. Part 2 reviews the design for packed SWS columns and presents a summary of potential issues that may be encountered in the operation of a SWS system.

**Packed tower design for SWSs.** Packed towers in SWS service are not as frequently encountered as trayed towers, but have been designed and operated successfully in units processing relatively clean water.[15] Packed SWSs may be used instead of trays for the following reasons:

- **Familiarity**—The facility may have experience with operating packed SWSs.
- **Pressure drop**—The possibility of achieving a lower pressure drop with a packed column may have benefits for some systems.
- **Cost**—Packed columns may be perceived to be less expensive than trayed columns, including the column and internals.
- **Wider operating window**—Packing may allow more turndown, which could be important for refineries that must run at low rates for a period or switch to crudes containing much less sulfur and nitrogen.
- **Equipment reuse**—A facility may be able to reuse a packed column from another process as the SWS, or to upgrade performance of an existing packed SWS by upgrading the internals (such as distributors and packing).
- **Size**—It may be easier to use packing for small or very small SWSs.

However, the major drawback to packed towers is that the packing can trap particulate matter as the sour water flows down the packed section. Over time, the packed sections of the tower can become fouled and maldistribution across the bed(s) of packing may result. Therefore, pretreatment of the sour water with the separation equipment—as described in Part 1—is very important. It is also critical that fouling-resistant distributors and packing be used in the stripper.

**Liquid distributors.** Distributors designed for fouling service are essential for successful operation of packed SWSs. Maldistribution of liquid in the top of the SWS will negatively impact the efficiency of the entire stripper. Redistribution by the packing will not be able to overcome any maldistribution from the distributor.

As such, there are trade-offs with liquid distributors that need to be considered to avoid the tendency to plug and foul, while also providing adequate distribution of liquid over the packing. General recommendations for liquid distributors in sour water service include:

- Using larger orifices to minimize fouling and plugging of orifices
- Reducing the number of drip points—generally an effect of using larger orifices, but not less than 5 points/$ft^2$
- Using orifices in the sides of the distributor walls and not at the bottom
- Maintaining levelness of the distributor in designs that use a gravity driving force.

Several different types of distributors that could be used in sour water service are described next. A vendor should be contacted to review the specific sour water application and to make a recommendation on the type of distributor most suitable for that service.

**Channel-type distributors.** An example of a channel-type distributor used in SWSs is shown in **FIG. 5.** The channel-type distributor has holes in the sides of the distribution channels. This type of distributor is plugging resistant, with generally good overall distribution. The holes in the distribution channels should be as large as possible, given the minimum drip point density allowed by the distributor design. All models have guides of some sort, such as plates and drip tubes. These guides are critical for good distribution. Channel wall designs are also illustrated.

**Weir riser pan distributors.** An example of a weir riser pan distributor is shown in **FIG. 6.** This type of distributor is used for smaller-diameter (12-in.–48-in.) towers in highly fouling service. The distributor is relatively inexpensive. As shown, the weirs serve as both liquid downcomers and vapor risers. A v-

notch allows for distribution of a large range of liquid flowrates. This type of distributor design is used with heavily contaminated liquids and highly fouling service. However, it does not provide distribution as effectively as some other designs. Although the figure shows a v-notch, a rectangular notch is preferred.

**Trough distributors.** Trough-style distributors, with notches in the trough wall, have improved fouling resistance and have been proven to be suitable for SWSs. A trough-style distributor will usually have the liquid feed into a parting box that distributes the liquid to individual troughs, and then liquid flows out of the individual troughs through vertical rectangular slots (or notches) cut into the side walls of the trough. **FIG. 7** illustrates a typical trough distributor with a single parting box.

The parting box feeds liquid to the troughs through windows cut into the parting box wall, so that the entire distributor is gravity fed. Notches in the distributor must be to mitigate fouling concerns, and this limits the efficiency of the distributor. The notches have a vertical rectangular slot with a V at the top for overflow. Installing the distributor on a level plane is critical to ensure that the distributor wets the packing below evenly—any poor liquid distribution in the distributor will negatively impact the efficiency of the entire packed bed.

**Spray nozzle distributors.** A spray nozzle distributor is shown in **FIG. 8.** Spray nozzle distributors have been proven in various severe fouling services in refinery units. Although opinions vary as to whether spray nozzle distributors should be used in the rectifying part of a SWS (as opposed to being used in a pumparound condenser, if present), this type of distributor is relatively familiar to refiners, which is why such a distributor is discussed here. To use this type of distributor, a higher-pressure liquid source is required. The design typically uses nozzles with

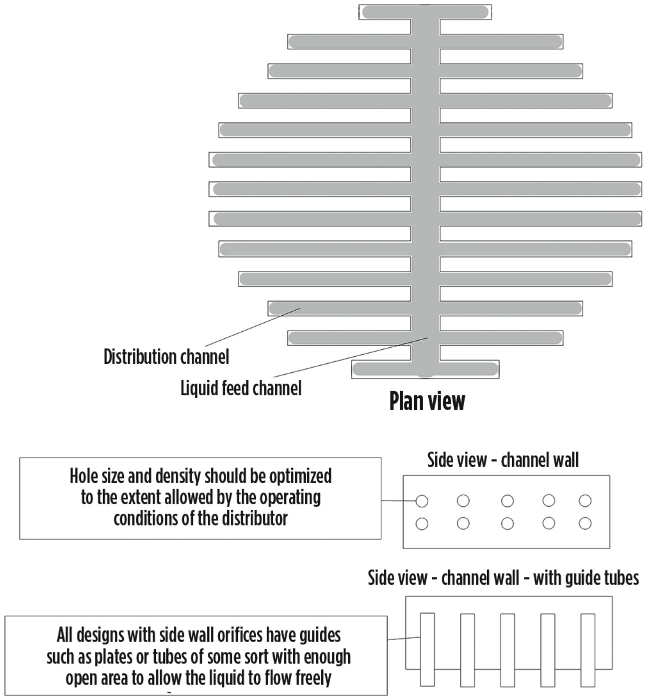


**FIG. 5.** An example of a channel-type distributor.

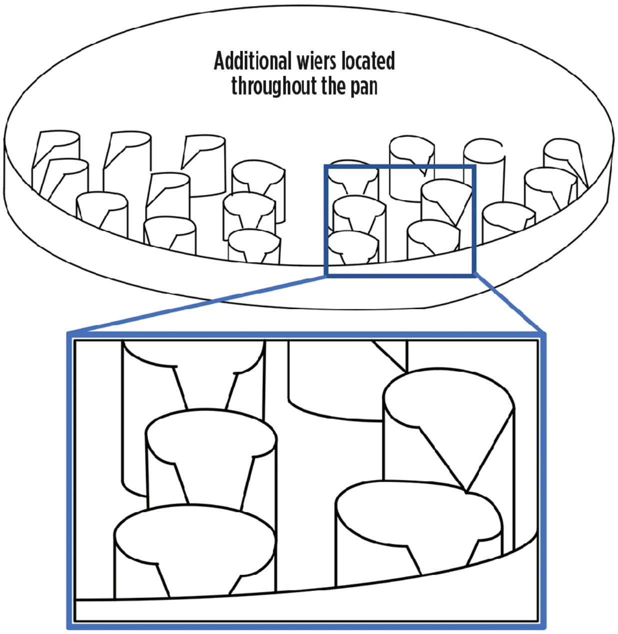


**FIG. 6.** An example of a weir riser pan distributor.

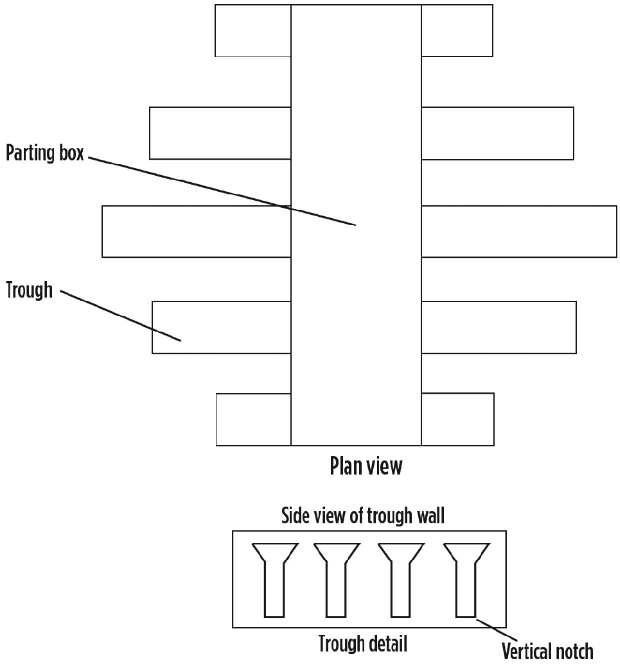


**FIG. 7.** Sketch of a trough-style distributor.

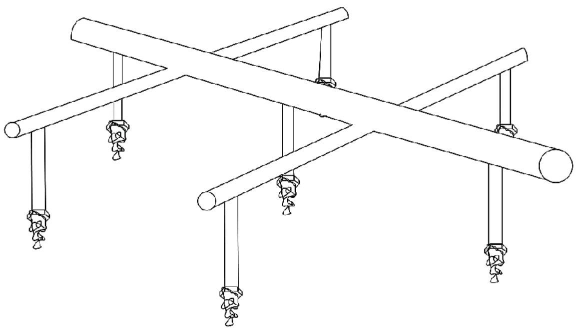

**FIG. 8.** An example of a spray header liquid distributor.

the maximum amount of free-passage available, and is often in a full cone spiral design (**FIG. 9**). While this design may have poorer distribution than some of the other distributors mentioned in this article, it can be more fouling resistant.

**Liquid distributor comparison.** A simplified example of characteristics of the liquid distributors described previously (based, in part, on information in literature) is shown in **TABLE 1.**[15] Specific vendor designs may vary but will generally have these relative characteristics.

## Packing height of an equivalent theoretical plate (HETP).

Packed tower design in SWSs will run into the same issue encountered in trayed towers when considering the efficiency of the trays. In this case, the efficiency of the separation in a packed tower is represented in some cases by the HETP. Like tray efficiency, the HETP is a chemical engineering factor that is not static from one separation to another or even from one packed bed to another in the same tower.

Packed towers in SWS service generally use dumped (random) packing. HETP values for these types of packing are available from many vendors. Published HETP values are not specific to SWS operations. If general vendor-published HETP values are used without considering the conditions that could be present in a SWS, the system will not likely work correctly or for long. The mass transfer limitations must be considered.

Rules of thumb for packing have been offered in discussions at prior industry conferences, and the authors' company has discussed these in conversations with several refinery subject matter experts (SMEs). A rule of thumb for 2-in., second- or third-generation packing is to use 2 ft of packing depth per actual tray. Given the previous rule of thumb for trays (three actual trays per a theoretical stage in SWS service), and assuming a tray spacing of 24 in., this rule of thumb makes the height of packing the same as the height of trays that would have been present—if trays had been chosen. If one assumes that third-generation packing has similar efficiency to second-generation packing, then the rule of thumb seems applicable to both generations. However, the rule of thumb may be overly conservative. For example, one should nominally see an increase in capacity, or an increase in efficiency, or possibly both, when going from second-generation packing to third-generation packing. Also, better efficiency may be possible if the sour water is as clean as possible (i.e., if good feed preparation steps have been used) and if the liquid distribution quality is as good as possible. One author reported going from 2-in., second-generation packing to 1.5-in., third-generation packing, and achieving a significant improvement in efficiency in the same bed height and capacity.[4]

**TABLE 2** shows HETP values from the literature and from three actual operating SWSs. The data shows relatively good agreement between the actual HETP and published HETP data for Source 2 (0.7–0.9 ratio) and Source 3 (1.1–1.3 ratio). However, the actual/experienced HETP for Source 1 was more than two times the published HETP. Several reasons may explain this. Discussions with the author for Source 3 indicated that a reasonably good distributor (in this case, meaning good balance of liquid distribution and low fouling tendency) was used in the stripper. Source 1 was known to have a poorer distributor type (not one of those mentioned previously). Source 1 also had other issues, including areas of plugged packing in sections of the tower. Furthermore, conditioning of the sour water feed may have been different between the sources—the literature discussing Source 3 mentions "minimal historical fouling and foaming issues"[4], so it may have had a cleaner sour water feed stream.

**TABLE 2** also shows the ratio of the SME-designed HETP to the vendor/published HETP. This factor ranged from 1.8–2.4, indicating that many operators choose to use a much larger HETP, conservatively increasing the packing bed requirements and the sizing of SWSs.

## Packing recommendations.

Overall, the following may be useful when considering using packing in SWS service:

1. Distributors should seek to balance a fouling-resistant design with good liquid distribution (adequate drip point density); trough-style liquid distributors with

**TABLE 1.** Comparison of liquid distributors for SWSs

| Parameter | Channel | Trough | Weir riser pan | Spray |
|---|---|---|---|---|
| Driving force | Gravity | Gravity | Gravity | Pressure |
| Tower size | Typically medium to large | Typically medium to large | Typically small | Any |
| Liquid distribution quality (of those listed) | Best | Lower | Lower | Lower |
| Propensity to plug | Low to medium | Low | Low | Low to medium (depending on nozzle) |
| Must be installed almost perfectly level | Yes | Yes | Yes | No |
| Requires precise nozzle aiming | No | No | No | Yes |

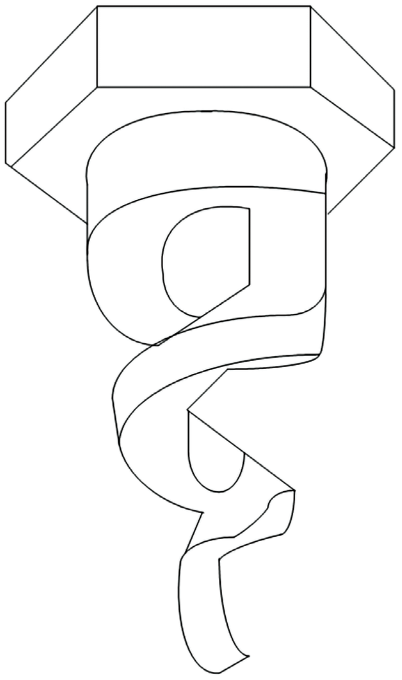

**FIG. 9.** Sketch of a full cone spray nozzle with maximum free passage.

**TABLE 2.** HETP comparison—Actual operation

| Parameter | Source 1 | Source 2[17] | Source 3[4] |
|---|---|---|---|
| Packing type | Small-diameter, ring-type packing | 1-in. ring | 1.5-in. specialty |
| Actual HETP, in. | More than 34 | 13 | 21 |
| Vendor/literature published HETP, in. | ~ 17 | 15, 17, 19 | 20, 16 |
| Ratio of actual HETP to published HETP | More than 2 | 0.9, 0.8, 0.7 | 1.1, 1.3 |
| Ratio of SME-designed HETP to published HETP | | ~ 1.8–2.4 | |

large liquid openings in the side walls of the trough (either rectangular-notched or round holes) have given good service in this application.

2. Packing should be an open design without small openings to minimize the potential for fouling of the packing. The major packing vendors offer such open packing, with some explicitly marketed for SWSs.
3. HETP values published in vendor literature, or correlations provided in literature, should not be used directly for estimating required packing bed depth. It is necessary to consider the potential for fouling and efficiency loss in the SWS packing. Although some SWSs have experienced actual HETPs approaching vendor/literature HETP values, it is suspected that those systems were fed sour water that was cleaner than is typically found, perhaps due to practices like those mentioned in the next point. Experience and good engineering judgment must be used in estimating HETP values that will be experienced in the end, which should include evaluation of the mass transfer.
4. Sour water feed conditioning systems are likely even more critical for a packed SWS than they are for a trayed SWS. Adequate three-phase separation in all separators in the process (even in the reflux drum, if installed) is recommended. Adequate settling time in the sour water surge tank is also recommended, in addition to the particulate filter and liquid coalescer.

**Issues encountered in operation.** A wide range of operating problems can occur in an SWS system. A select few are discussed in the following subsections.

**Fouling of SWS internals.** SWS internals can foul from many different materials. Corrosion products can accumulate on the tray or in the packing and cause fouling.

Even with all the preventive measures discussed in Part 1, hydrocarbons and solids may still enter and foul the SWS internals. It has also been reported in literature[6] that, because of the high vapor pressure of water in the stripper, volatile hydrocarbons will evaporate with the overhead gas. As a result, removal of lighter hydrocarbons may make heavier hydrocarbons less soluble in the water and too viscous to flow properly at the SWS temperature.[6] The heavy hydrocarbons collect, along with corrosion particulates and other solids, to form a fouling layer on trays or packing.[6]

If the SWS is underperforming and other more routine process checks on the system have not identified a cause, then a gamma scan can be conducted to determine if the internals of the tower are damaged or severely fouled. A gamma scan generates a density profile of the column that can be used to identify the integrity of internals and column operating conditions. Scans have shown columns where entire packed sections were missing or lower than expected. Maldistribution of liquid in the column can also be demonstrated via the scans. Maldistribution of liquid in the column will reduce the efficiency of the packing. Issues with the integrity of tray towers can also be identified.

**Maintenance and monitoring requirements.** Routine maintenance and cleaning of equipment may be prudent to remove fouling and particulate, and to improve the run time of the SWS system. For example, exchangers with bypasses can be periodically cleaned online. Exchangers used in other processes that transfer heat between a sour water stream and a hydrocarbon stream should be inspected for leaks to minimize the potential for sending hydrocarbon-contaminated water to the SWS system.

The SWS tower could also be washed occasionally if there is enough storage for the sour water at the plant. Weak acid and base washes can remove scale, and detergent washes can remove hydrocarbons.[7]

The hydrocarbon and liquid levels in the flash drum and surge tank should be routinely visually checked to ensure that they are at the proper heights and that hydrocarbons are not entering the stripper. Level controls and interface level controllers in the flash drum and surge tank should be inspected on a routine basis to make sure they are working appropriately.

The SWS overhead lines should be periodically examined for cold areas [less than approximately 82.2°C (180°F)] to prevent salts from depositing. The overhead lines need to be steam traced and insulated or steam jacketed.

Process instrumentation should be routinely checked for accurate readings to aid in diagnosing potential SWS problems. The column differential pressure, overhead temperature and process water flow are important parameters to monitor.[7]

In some cases, chemical agents (such as dispersants, scale/corrosion inhibitors and cleaning solutions) may be able to help control or remove fouling from the residue of hydrocarbons, salts and corrosion byproducts.

Solid and liquid material from the filter, liquid coalescer or other equipment could be analyzed to determine the type and source of fouling. Routine samples of the sour water and stripped water should also be taken to help identify issues in performance.

**Salt solids formation.** The formation of salt solids is another concern in SWSs. For example, in SWSs that remove $H_2S$ and $NH_3$, ammonium bisulfide ($NH_4HS$) solids may form in the overheads line. When the acid gas condenses, the reflux water may contain a high concentration of $NH_4HS$, which can lead to corrosion and salt solids formation. An SWS with reflux usually has higher concentrations of $NH_4HS$, but pumparound systems can also be impacted by this type of corrosion.[7] Corrosion increases with increasing $NH_4HS$ concentration and velocity. Carbon steel is often acceptable when the $NH_4HS$ concentration is less than approximately 2 wt%. Carbon steel is marginal when the concentration is between 2 wt% and 8 wt%. Above 8 wt%, carbon steel is generally viewed as unacceptable, and stainless steel or other higher alloys may be required.[18] The overhead temperature is generally kept at or above 82.2°C (180°F) to avoid the formation of $NH_4HS$ solids that can plug lines and equipment.[6,19]

Other salt solids can be present in SWS systems, as well. Salt solids can form if the water feed is hard (i.e., if it contains a significant amount of calcium and magnesium). Among other causes, salt solids may occur if low-quality wash water is used in the equipment generating the sour water.

Ammonium carbamate ($NH_4CO_2NH_2$), ammonium bicarbonate ($NH_4HCO_3$) and ammonium carbonate [$(NH_4)_2CO_3$] solids can form if carbon dioxide ($CO_2$) is present. $(NH_4)_2CO_3$ and $NH_4HCO_3$ will sublime from the stripper overhead gas at temperatures of 54.4°C–75°C (137°F–167°F).[7] The deposition temperature depends on the partial pressures of the acid gas components ($NH_3$, $H_2S$ and $CO_2$) and $H_2O$ in the stream. Deposition curves exist in the literature for many of these salts. It is considered best practice to operate the SWS a safe margin above the estimated sublimation/deposition temperature.

**Takeaway.** SWS is a demanding process in a refinery or gas treating facility. The sour water will contain a multitude of contaminants in addition to the $NH_3$ and $H_2S$ stripped out of the water in the process. These contaminants make reliable operation of the SWS a challenge. Part 1 described the equipment upstream of the SWS that reduces fouling and foaming issues in the stripper tower, including the sour water flash drum, surge/storage tank and solids filtration/coalescing filters. Part 1 also presented design guidance to account for foaming in the stripper and optimum feed location for the sour water. The design features for trayed columns were also reviewed in detail.

Part 2 focused on the internals specified for SWSs with a packed tower design. The types of liquid distributors used in SWS service were discussed, along with a comparison of HETP values from vendors, actual operation and SMEs. Other packing recommendations were provided, and operating issues encountered in SWS systems were also discussed. Even with all the upstream equipment designed to remove contaminants, corrosion products can still accumulate on the tower internals and cause fouling. Salt solids formation—such as $NH_4HS$, $NH_4CO_2NH_2$ and $NH_4HCO_3$ solids—can also occur at cool spots in the overheads line, requiring proper temperature management. Regular maintenance and monitoring can improve SWS performance and extend the run time for the system, which will benefit overall refinery operations. HP

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# Getting the most out of your process burner tiles

It is often incorrectly assumed that the burner tile is simply a piece of refractory or firebrick. Nothing could be further from the truth. The burner tile is the "heart" of the burner. Tile that has not been poured correctly, installed correctly or that has not been properly maintained can have a serious impact on burner performance.

Burner tiles, sometimes called quarls, are an important component in burners.[1] Tiles are typically made out of some type of ceramic and serve many different purposes, including:

- Protecting the metal components inside the burner from the heat in the combustion chamber
- Shaping the flame
- Radiating heat: these burners are commonly called radiant wall burners (FIG. 1) as they are normally mounted on the side wall of a heater.[2]

Many potential issues are related to the tile. For example, if the ceramic used to make the tile is improperly cured, the tile may be damaged during startup. If the burner tile is improperly installed, this can adversely affect the operation and performance of the burner, including the emissions. If a burner tile is damaged, burner performance may suffer.

The furnace temperature and composition of the combustion products can affect the materials used in the tile. Burners fired downward, for example in a downfired reformer, require special mounting considerations. If the tile is oversized, the burner may not operate correctly because the air-fuel mixing and flame shaping may be adversely affected. If the tile is undersized, the pressure drop may be too high and the firing capacity reduced.

These elements are discussed here in detail, as well as general fabrication, installation and troubleshooting issues related to burner tiles.

## UNDERSTANDING THE SIX "M"S

Six "M"s generally can be used to describe general burner design principles[3] that can be applied specifically to the tile.

**Meter the fuel and air.** The tiles help meter air flow. The tile throat is sized to allow a certain amount of air into the combustion reaction at a given pressure drop (FIG. 2). If the throat is too small or too large, the burner will not operate with the required amount of excess air. If the tile is too small, then not enough air can be pulled through a natural draft burner, or the pressure drop and power requirements will increase in a forced draft burner.

It may be tempting to oversize the tile throat in case the burner firing rate needs to be increased at some point in the future. However, an oversized tile reduces the turndown in a natural draft burner. For example, if the tile is sized for a 10 MMBtu/hr-burner capacity and has a 5:1 turndown, then its lowest design firing rate would be 2 MMBtu/hr. If the actual operating conditions today require only 5 MMBtu/hr, then the effective turndown is 5:2 or 2.5:1. Oversizing the tile for a possible firing rate of 10 MMBtu/hr is also detrimental to burner performance because the burner is designed to work optimally at 10 MMBtu/hr. If it is fired at half rate, it may not get as low emissions as it would have at the full design rate, for example.

**Mix the fuel and air.** Proper mixing of the fuel and air is critical to process burner operation. Burner tiles are often used to help mix the fuel and air using a variety of techniques. One such technique is to have ledges inside and/or at the top of the tile. These ledges are bluff bodies that promote mixing. For example, a ledge at the top of a tile can be used to cause the combustion air inside the tile and the fuel on the outside (from secondary or staged fuel injectors) to form vortices on the ledge to promote mixing. It may be desirable to delay mixing until the end of the burner to prevent flashback. It may also be desirable to reduce nitrogen oxide ($NO_x$) emissions due to the delayed combustion that minimizes hot spots in the flame.

Holes through the side of some tiles help mix fuel and air. Some burner designs have fuel injectors or tips outside

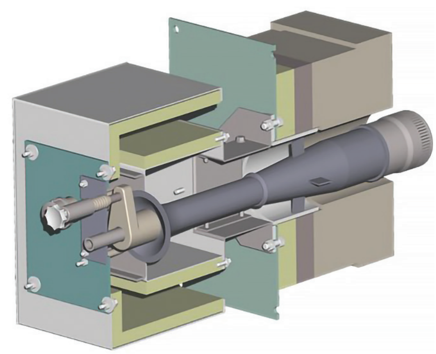

FIG. 1. A cutaway of a radiant wall burner.

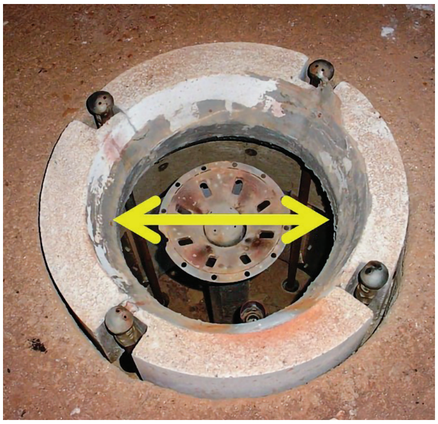

FIG. 2. A burner tile throat, which creates an orifice to meter the combustion air flow.

the tile. Those tips may inject fuel up along the outside of the tile to delay mixing for $NO_x$ reduction,[4] and inject some fuel through holes in the side of the tile into the main or primary flame zone. The holes may be straight through the tile or they may be at an angle to impart some swirl to the flame. In either case, they are used to help mix the fuel and air. A burner tile can have secondary air outlets built into the tile. The primary air comes through the middle of the burner throat. The purpose of the secondary air is for staging to reduce $NO_x$ emissions.

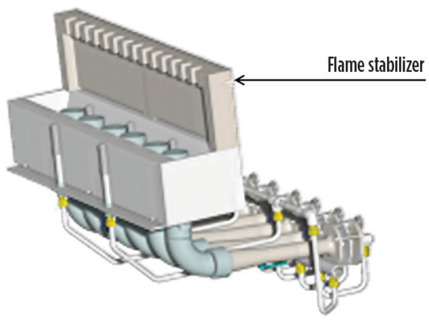


**FIG. 3.** A burner tile with "teeth" for flame stabilization.

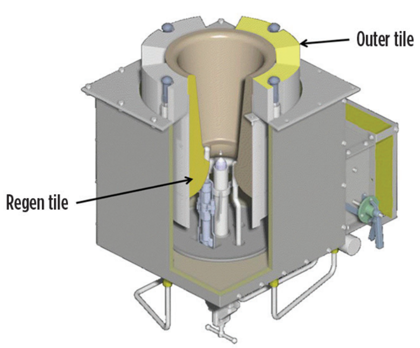


**FIG. 4.** Example of an inner regen tile for a combination burner capable of firing on gas or liquid fuel.

**Maintain ignition.** An important safety consideration in any combustion system is to maintain ignition when fuel is flowing. Otherwise, that fuel may ignite somewhere else, which can prove to be dangerous.

The tile ledges may be designed to help anchor the flame and sustain combustion, as previously described for mixing. Many process burners are diffusion or raw gas (also known as nozzle-mix) burners, which means the fuel and air are mixed at the burner outlet. This is in contrast to premix burners, which mix some or all of the fuel and air inside the metal parts of the burner. While some advantages of premix burners exist, a significant disadvantage is the possibility of flashback. The tile used in raw gas burners is specifically designed to mix the fuel and air at the outlet to anchor the flame throughout the design firing rate range.

The visible flame may start a short distance away from the tile, but it should not be a very long distance, regardless of the firing rate. If the gap between the end of the tile and the start of the visible flame is too long, then the condition is referred to as a "lifted flame." Lifted flames are undesirable, as they are much more susceptible to lifting off completely, causing the flame to be extinguished. If fuel continues to flow to that burner and components in the heater—such as the burner tile or heater walls—are above the autoignition temperature for the fuel/air mixture, then a re-ignition of the mixture is likely, which can lead to an explosion. An obvious sign of an unstable flame is pulsing or huffing (where the flame is bouncing up and down). This is a very dangerous condition that must be corrected immediately.

**FIG. 3** shows an example of a burner tile with a specially-designed edge that has the appearance of teeth, for flame stabilization.

Another important function of the tile when firing liquid fuels is to produce a hot chamber where the liquid fuel is atomized. The higher temperatures help promote and maintain liquid vaporization. Some burners are designed to recirculate hot combustion products inside the tile to aid in the vaporization. These tiles are sometimes called "regen tiles," which is short for "regenerative tiles" (**FIG. 4**).

**Mold the flame.** The shape of the tile outlet helps determine the cross-section of the flame: round tiles make round flames and rectangular tiles make rectangular flames (**FIG. 5**). These are by far the two most popular tile cross-sections for process burners.[5]

**Minimize emissions.** In some burner designs, the tile helps reduce $NO_x$ (e.g., a burner tile where secondary fuel tips are in the tile wall). The purpose is to stage the secondary fuel into the flame to reduce hot spots, reducing $NO_x$ emissions.

**FIG. 6** shows an advanced burner tile designed to help minimize $NO_x$ formation. The tile has two different alternating slopes with dividers in between. Primary tips inject some fuel through the tile into the primary flame zone inside the tile. Those primary tips also inject some fuel along the outside of the tile. The staged tips only inject fuel along the outside of the tile. The overall purpose of the design is to minimize the amount of fuel in the primary zone and maximize the amount of fuel in the secondary flame zone. This fuel staging reduces hot spots in the flame, minimizing $NO_x$ formation.

Another important aspect of this tile design are the holes going through the

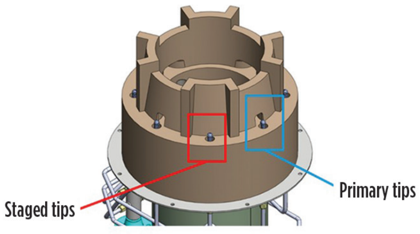


**FIG. 6.** Advanced burner tile design to help minimize $NO_x$ formation.

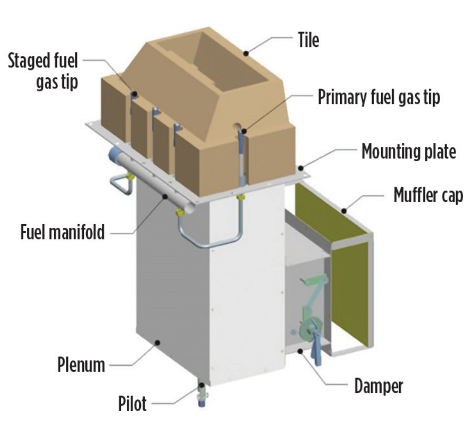


**FIG. 5.** Burner with a rectangular tile.

side of the tile. In addition to the high-speed fuel from the primary tips flowing through those holes, furnace gases are entrained by the fuel and flow into the primary flame zone, as well. That entrainment of furnace gases helps to homogenize the flame temperature and reduce flame hot spots, which helps minimize $NO_x$. The size of those holes is important. If the holes are too small, less furnace gas is entrained and more $NO_x$ is produced; if the holes are too big, too much furnace gas could be entrained, which could cause flame instability.

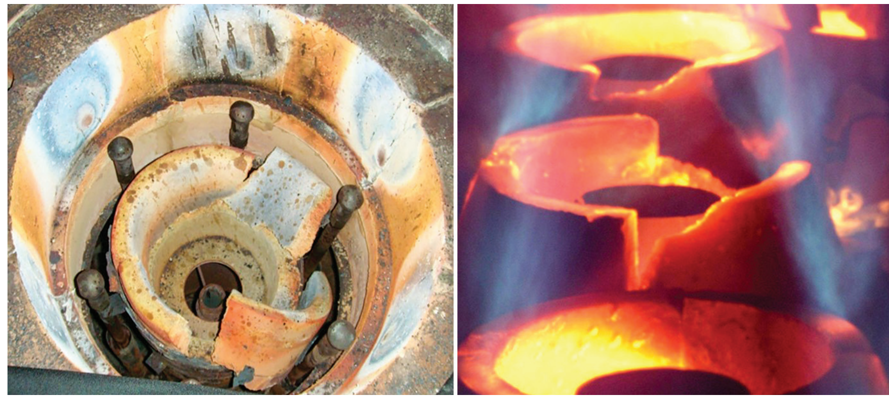

**FIG. 7.** Damaged (left) regen tile and (right) burner tiles.

**Minimize costs.** While many tile shapes could be made, round and rectangular are by far the most common shapes for process burners. They are typically the cheapest to make because the tile molds are simpler to build. Those tile shapes are also usually stronger because they have fewer sharp changes in direction that produce corners, which are more likely to crack.

**Tile fabrication, installation and maintenance.** Proper tile fabrication is critical to burner performance. Obviously, the shape must be correct, but the actual fabrication process is important, as well. For example, if the ceramic is made from a water-based slurry, it must be slowly dried to prevent steam from rapidly forming inside the mold, which could cause the ceramic to spall apart. If the tile is improperly dried prior to installation in a heater, a rapid startup could also cause steam formation in the tile to spall the refractory. There are chemically-base ceramic materials that do not need to be slowly dried because of the lack of water in the ceramic.

Burner tiles will not operate properly if they are installed incorrectly.[6] Some burner tiles consist of four segments or sections that make it easier to lift the tile into position without the use of a forklift. It is important that enough—but not too much—mortar is applied between the segments. The burner drawings and manual should be followed to make sure the proper tile throat area is maintained. Tiles must be properly aligned to allow the design combustion air flowrate.[7]

Tile maintenance is relatively simple: ensure that any damaged tiles are repaired or replaced. **FIG. 7** shows some tiles that have been damaged beyond repair. **FIG. 8** shows a tile that has been coated with catalyst. That coating reduces the burner outlet, which reduces the air flow through the tile for a given draft level. Some cracks in the tile are evident, as well.

**FIG. 8.** Catalyst buildup on a regen tile.

**Troubleshooting and takeaways.** Burner tiles usually degrade over a period of time, although they can become damaged suddenly, as well. For example, refractory falling out of the roof onto a tile can cause damage. Operators should examine burner tiles on a regular basis during heater operation. Burners can be operated with damaged tiles, as long as the damage is not too severe. A much closer inspection should be made during maintenance turnarounds. Typically, turnarounds are somewhat infrequent, so any damaged tiles should be repaired or replaced during turnarounds as another chance may not be available for some time.

The burner tile is an important and integral part of a process burner that may impact the fuel/air mixing, burner stability, flame shape and pollution emissions. It must be properly manufactured, installed and maintained for optimum performance. A variety of possible tile problems exist, such as cracking, buildup, chemical attack and improper installation. Most of these are easy to detect and should be corrected, as appropriate. HP


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# The use of tube inserts in fired heaters

In 1896, Whitham[1] reported the successful use of twisted tapes (originally called retarders, and now also called turbulators) to increase heat transfer in boiler fire tubes—their effectiveness in increasing heat transfer is well known. Engineers usually use twisted tapes to improve heat transfer in laminar flows, but academics and industry have extensively studied their use in turbulent flows.[2]

Refining and petrochemical plants commonly use fired heaters where the process requires high-intensity heat. State-of-the-art fired heaters are some of the most fuel-efficient devices in use, with efficiencies over 92%. It is not readily apparent how the use of turbulators might benefit such highly efficient systems. However, using the knowledge that the process flow in many fired heaters at least partially vaporizes, one can show that properly applied tube inserts that outwardly resemble turbulators can improve fired heater performance.

**Fired heaters, heat transfer and heat flux uniformity.** **FIG. 1** shows a typical fired heater. Burners fire into a "radiant section," transferring heat from the flue gas to the process fluid, which flows through pipes commonly called "tubes." The process fluid typically receives 60%–70% of the heat within the radiant section. The flue gas then flows into the "convection section," which transfers 15%–20% of the remaining heat to the process fluid. The principal mode of heat transfer in the radiant section is thermal irradiation, while the principal mode of heat transfer in the convection section is convective.

The process flow in practical fired heaters is turbulent, with a Reynolds number on the order of $10^6$. Most of the heat transferred to the process occurs within the radiant section. The convection section compensates for small reductions in radiant section heat transfer due to fouling and non-ideal flames; a higher radiant section exit temperature results in more heat transfer within the convection section. Fired heater designers have had little incentive to increase heat transfer intensity using turbulators. The process flow is turbulent, resulting in a high tube-side convection heat transfer coefficient, and the convection section design already results in the desired flue gas exit temperature. However, the benefit of in-tube mixing enhancement goes far beyond the benefits of increased heat transfer.

Reformers and pyrolysis heaters fall among a class of heaters where intentional and valuable reactions take place inside the heater tubes. In some reactor charge heater tubes, undesirable chemical reactions cause in-heater feed conversion, which reduces the ultimate yield. Other heaters produce unintended reactions that reduce the value of the product. Heaters used in certain services, such as crude distillation, vacuum distillation and delayed coking, have both unintended reactions and phase change within the heater tubes. Similar devices to heaters, such as once-through steam generators (OTSGs), do not have chemical reactions within the tubes but do exhibit phase change. In all cases, not only is the total absorbed heat important, but also the location of the absorbed heat.

To see why the variation in temperature along the outside coil surface is critical in heaters with multiphase process flow, consider the idealized graph of heat transfer coefficient vs. temperature difference between wall and fluid, illustrated in **FIG. 2.** Beneath the chart is a representative picture of the liquid/vapor composition within the process coil corresponding to the heat transfer coefficient. The highest temperature at any point in the flow occurs at the boundary between the process coil wall and the process flow. When the combustion heats the process flow, vapor first forms at the interior wall of the pipe. Gases have a significantly lower overall convection heat transfer coefficient when compared to liquids, so the process flow transfers less heat away from the pipe wall. In this way, adverse feedback ensues wherein the high wall temperature begets more boiling that, in turn, reduces the inside convection heat transfer coefficient, thereby increasing the wall temperature and the resulting boiling. In this way, "hotspots" can form on the heater tubes given an initial slight difference in heat transfer.

Gravity further exacerbates phase nonuniformity inside horizontal sections of process coils. Gravity pulls liquid—being denser than vapor—to the bottom of the

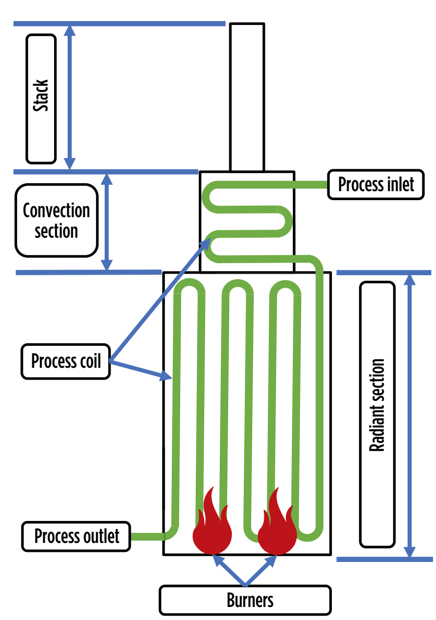


**FIG. 1.** A typical fired heater. Burners fire into a radiant section, generating hot flue gas. Radiation transfers 60%–70% of the heat in this section, after which the flue gas flows through the convection section where the process absorbs 15%–20% of the remaining heat.

pipe, resulting in a strong tendency toward stratified flow with resulting higher temperature on the upper pipe surface.

The major sources of temperature non-uniformity result from an unavoidable combination of geometry and physics. The burner flames and hot flue gas transfer more radiant heat to the flame-facing surface of the coil. Heaters with burners mounted on one or both sides of the coil exhibit this non-uniformity. The measure of this non-uniformity is known as the circumferential flux factor and has been well-characterized. As combustion releases heat from the burner flame, an additional longitudinal flux factor exists.

Engineers have been unable to characterize the longitudinal flux factor in a general sense because it is heater and burner dependent. **FIG. 3** shows an example of single-fired tubes with a high-longitudinal variation in flux and the resulting peak-to-average flux ratio of 2.3. This flux ratio translates to a 16.1°C (61°F) higher film temperature and a 37.8°C (100°F) higher tube metal temperature in example calculations.

Variation in heat transfer to the process fluid comes from both within and without the process coil. So, what is to be done? Variation in the physical properties of the fluid that are driven by vaporization inside the coil plays a strong role in the local tube metal temperature. It is also apparent that physics and geometry of the flue gas side drive non-uniformity from the outside the coil. A common method of increasing homogeneity in a process is to increase the mixing of the constituents. One can show through simulation that properly designed tube inserts can improve process homogeneity. This improvement can, in turn, be used to reduce tube metal temperatures, increase yield and increase heater capacity.

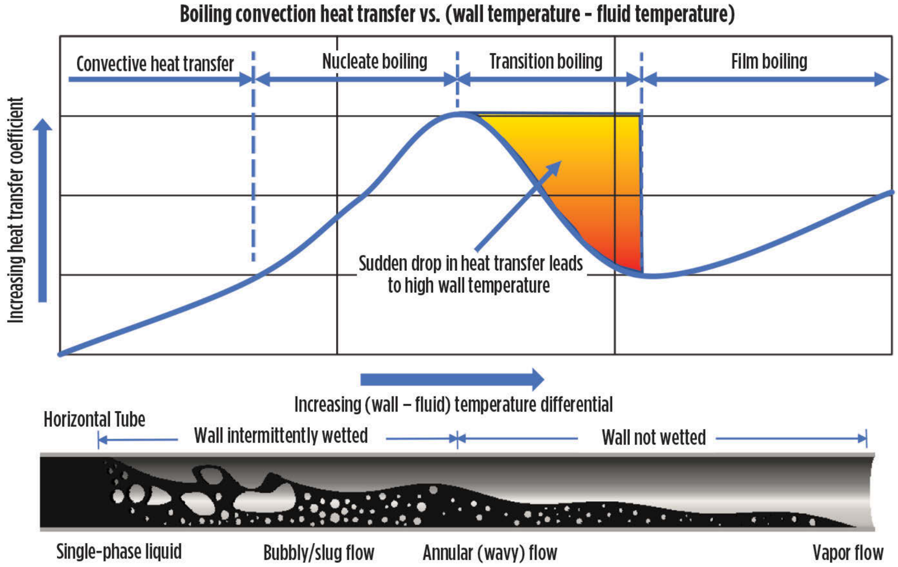


**FIG. 2.** An example heat transfer coefficient vs. temperature differential for a multiphase horizontal pipe and flow regime.

**Simulation results: Improved area goodness factor for tube inserts.** Increasing the mixing inside a process coil comes at a cost: the pressure drop through the system also increases. The process fluid diffusion inside the heater tube can be estimated by using heat transfer as a surrogate measure. To measure the relative effectiveness of various inserts, step-change designs were simulated in combination with automated optimization for thousands of combinations of geometric parameters. The relative merit of each design was judged by comparing the area goodness factor, or the Colburn factor ($j$) divided by the Fanning friction factor ($f$).

The final optimized design has a 30% increase in area goodness factor vs. traditional twisted tapes. The practical implication is that by using an optimized design, more process mixing with lower pressure drop can be achieved. A key parameter in twisted tape design is the twist-pitch, or the number of pipe diameters required for the twisted tape to make a helical revolution within a length of the tape. **FIG. 4** shows a comparison in the simulated area goodness factor for both traditional twisted tapes and the optimized design vs. increasing twist pitch.

**Simulation results: Multiphase flow through a horizontal return.** Simulations comparing the optimized design to an empty tube demonstrate that the in-coil mixing translates to increased process homogeneity for multiphase flows. **FIG. 5** shows the predicted convection heat transfer coefficient over the entire tube surface and liquid volume frac-

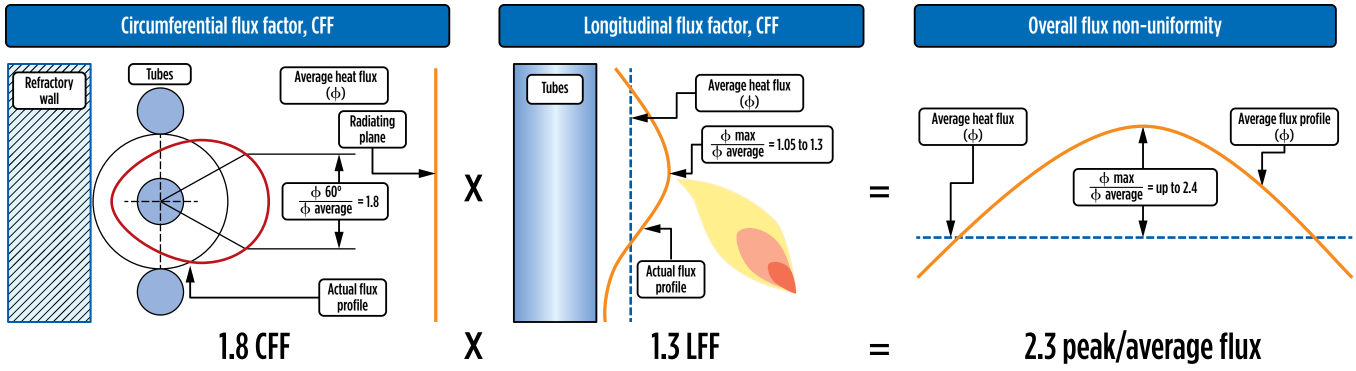


**FIG. 3.** Flux factors used in heater tube temperature calculations.

tion at the inlet and outlet of the tube for a simulation with 80% liquid and 20% vapor by volume. There are inserts both before and after the return, but in this case, the flow requires one straight section of tube to establish the mixing motion provided by the insert. With the tube insert in place, the convection heat transfer coefficient is more uniform when compared to the empty tube.

Near the inlet, at Point A, gravity stratifies both flows. At Point B, the return has temporarily changed the stratified flow to annular flow, producing a more uniform heat transfer coefficient at the tube surface for both cases. At Point C, the stratified flow returns in the empty tube, but the liquid remains adhered to the tube wall when using the insert.

On the outlet leg of the coil section, the tube insert increases the area-weighted average heat transfer coefficient by 20% when one compares the empty tube. More importantly, the minimum heat transfer coefficient over the same section of the coil using inserts is 50 times higher than the minimum heat transfer coefficient of the same section of the empty tube. Tube failures occur at specific, often initially small, locations. The use of the insert eliminates the point of minimum heat transfer coefficient where this failure would likely occur. By limiting the use of inserts to the tubes that are prone to failure or that need the most process improvement, their benefit can be maximized while reducing the additional pressure drop.

**Estimates of economic impact.** The variability in causes of shutdowns will impact any economic assessment for the use of in-tube inserts. If a plant shuts down often due to tube failures, the value for preventing those failures is far greater than incremental increases in run length. However, the value of using the inserts for normal operation without special cause failures can be estimated.

To evaluate the effect on run length, a model was created for a coker unit furnace. The computational fluid dynamics (CFD) model of the coil included submodels for phase change of the vacuum residual oil, as well as the propensity for coking based on the work of Ebert and Panchal.[3] The predicted coking rate and daily temperature rise for coils with the insert in place is three times less than of that of an empty tube. The predicted run length of the furnace is increased by 35%. Using validated assumptions for the coker spread, shutdown days and current energy costs, the 2-yr net value addition of inserts is $1.96 MM for a typical 30,000-bpd coker unit.

**Practical considerations and takeaway.** The use of tube inserts does raise the practical concern of how to remove the inserts when the coil requires pigging. A mechanical solution has been developed by the author's company that uses flanged connections outside the heat-affected zone. This solution removes the potential for leakage of the process fluid, which is a problem for traditional plug headers. Physical testing was performed to ensure that the insert can be removed from a tube even if coke or scale has completely seized it in place.

Tube inserts are a well-proven method to increase heat transfer. They also increase process homogeneity. The proper application of in-tube mixing should increase coil longevity, run length and unit profitability. These improvements should be more pronounced for processes with in-tube phase change. Properly optimized mixing elements provide both relatively low pressure drop and the increased mixing required to bring heaters to a new level of performance. HP


### ACKNOWLEDGEMENT

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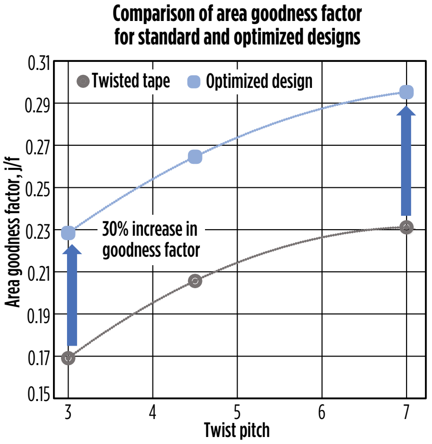


**FIG. 4.** A comparison of optimized tube insert design to a traditional twisted tape shows an increase in goodness factor of 30%.

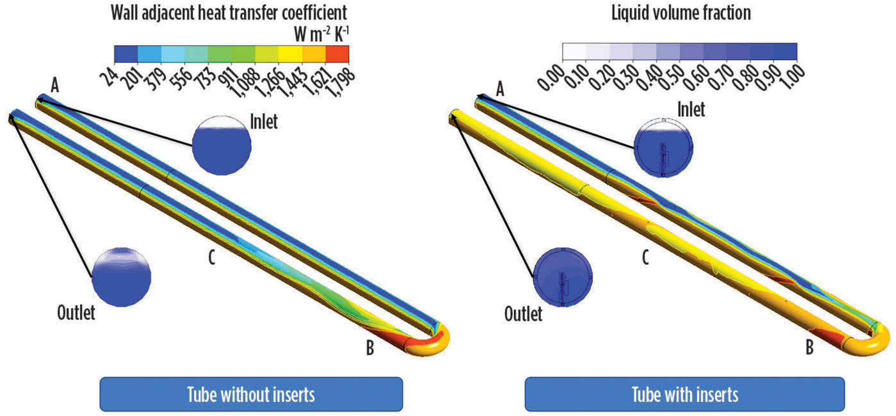


**FIG. 5.** Comparison of convection heat transfer coefficient and liquid volume fraction for a tube without (left) and with (right) tube inserts.

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# Optimizing failure data analysis

A critical objective of engineers who design an automatic protection system is to lower the false trip rate and to lower a key metric called average probability of failure on demand (PFDavg). The PFDavg metric represents the chance that the automatic protection will not work when needed. The false trip rate and PFDavg are based on several variables, including the failure rate of all devices used for each safety instrumented function (SIF). These devices are classified into three groups:

1. The sensor assembly, which detects a dangerous condition
2. The logic solver, which determines when to initiate the protection
3. The final element, which does the protection work, and is often a remote actuated valve that opens or closes.

Statistical analysis of more than 80,000 SIF designs shows that nearly 70% of the PFDavg comes from the final element (FIG. 1). Many engineers are looking for help from manufacturers to reduce this percentage and make their plants safer. Some manufacturers have had their products IEC 61508 certified for functional safety applications. While this helps, plant operators and engineers recognize that utilizing IEC 61508-certified products does not guarantee the reliability of the final element assembly. Many field-failure reports show root causes due to inadequate torque matching, insufficient assembly testing, manufacturing errors or application mismatching. More needs to be done to apply the lifecycle engineering rigor required by IEC 61508 to the entire final element supply chain.

**Pre-engineered final element assemblies.** Recently, organizations (such as WIB—The Process Automation Users' Association) have undertaken efforts to address this challenge by developing a new recommended practice (S 2812-X-1), which is targeted specifically at quarter-turn emergency isolation valve assemblies. This new recommended practice—titled "Actuated Valve Assembly"—significantly increases the rigor associated with the engineering, design, testing and documentation of complete final element assemblies. To satisfy the S 2812-X-1 recommended practice, those companies that supply final element assemblies must have strong engineering processes that allow the complete assembly to be engineered, designed and tested as a pre-engineered product. This added rigor has the potential to reduce the failure rates that contribute to false trips and the PFDavg.

To provide verification and structure, IEC 61508 certification can be applied to the assembly. To address this need, the authors' company has developed a certification scheme[a] that assesses the complete final element assembly. Several standards have been leveraged in the development of the certification scheme from organizations such as WIB, the International Organization for Standardization (ISO), the Instrument Society of America (ISA), European Norms (EN), the American Petroleum Institute (API) and the American Society of Mechanical Engineers (ASME). The resulting requirements from these relevant standards define a certification scheme that will reduce engineering errors and omissions, characterize lifetime torque, match actuator output, provide complete design testing and verify manufacturing quality. The program also requires a user document called a "safety manual" where safety design data—including application limitations, predicted failure rates, maintenance procedures

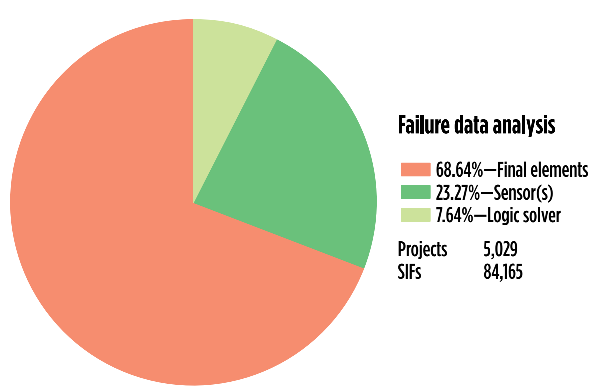



**FIG. 1.** Average PFD contribution results from more than 80,000 SIF designs. Source: exida exSILentia usage study, November 2019.

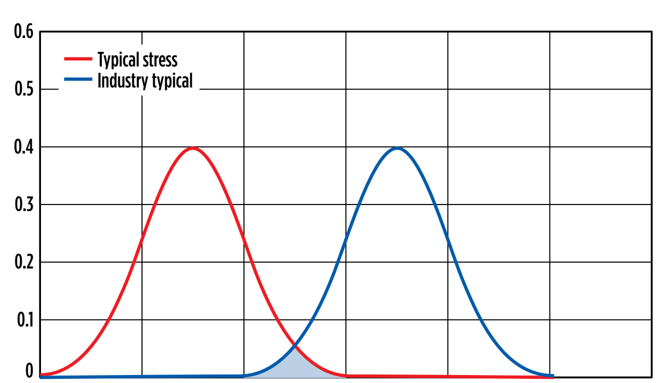



**FIG. 2.** A stress vs. strength distribution model.

and effective proof test procedures—are provided. The objective of the authors' company's certification program is to achieve fewer false trips and higher safety by increasing the strength of the assembly and reducing operational stress.

**Stress vs. strength.** An examination via the stress vs. strength distribution model illustrates this concept. FIG. 2 shows a case where stress and strength are normally distributed. A typical operational stress curve is shown in red, while the industry's typical strength is shown in blue. When stresses are greater than strength, a failure occurs. Failures represented by the gray area are known as the standardized interference distribution.

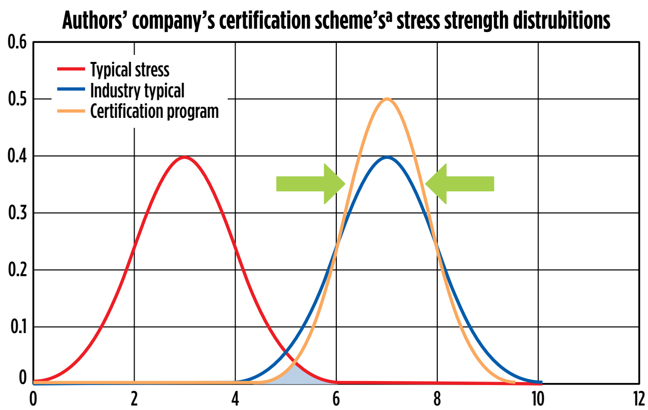



FIG. 3. Reduced strength variation in a final element assembly.

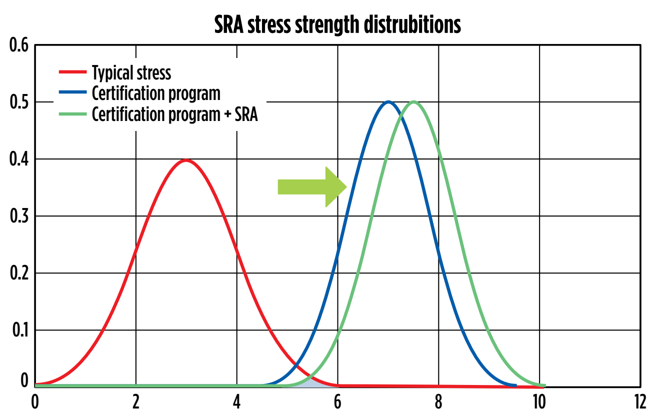



FIG. 4. Increased design/process strength.

Quality programs have shown that, when engineering design and manufacturing quality are improved, variation is decreased. FIG. 3 shows the impact of reduced variation in the strength of a population of products. The result is that the failure rate is improved. The authors' company's certification program requires a rigorous engineering, testing and manufacturing process. Candidate remote actuated valve assemblies are audited to ensure that the procedures are in place and that these procedures are being followed.

Additional strength can be added when manufacturers innovate and design a product to be uniquely fit for purpose (FIG. 4). When this can be documented, the strength curve is improved to reflect the increased design strength. Moving this curve again narrows the standardized interference distribution (gray area), further improving the predictive failure rate.

The authors' company's certification scheme includes a deep, detailed review of the design parameters to accurately estimate design strength. When a manufacturer has innovative designs with special coatings, surface finishes and strong strength margins, credit is given (FIG. 5). When weaknesses in design are removed from the devices and assemblies, strength is added, which results in a reduction of failure rates, along with a decrease in operational costs and downtime.

Increasing the strength is only a part of the story. By implementing an application review as part of the engineering process, the number of application mismatches is reduced. A stress vs. strength distribution model shows this as a reduction in stress variation. What does this really mean? It means that the chances of failure are getting smaller.

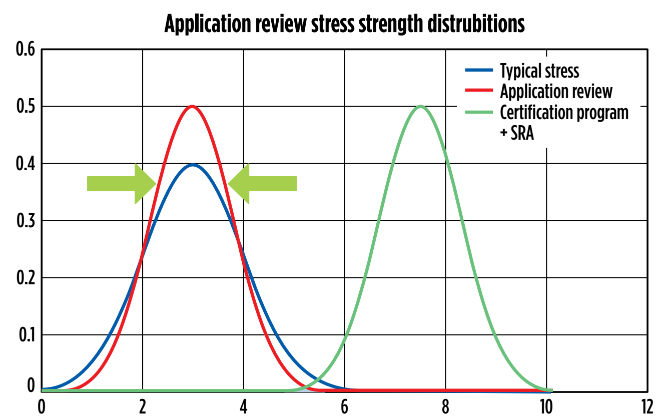
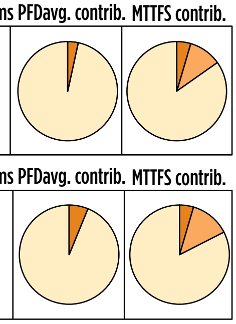



FIG. 5. Reduction in stress variation.



| | | Required | Achieved | | RRF | PFDavg. | MTTFS | SIL PFDavg. | SIL AC | SIL SC | Resp. time, ms | PFDavg. contrib. | MTTFS contrib. |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| SIF 1 traditional assembly | SIL | 2 | 1 | SENS | 2338.3 | 4.28E-04 | 3174.94 | 1 | 3 | N/A | 3000 | | |
| | | | | LS | 24375.5 | 4.10E-05 | 1206.33 | | 3 | N/A | | | |
| | RRF | 100 | 80.3 | FE | 83.4 | 1.20E-02 | 156.35 | | 2 | N/A | | | |
| | | | | SIF | 80.3 | 1.24E-02 | 132.63 | | 2 | N/A | | | |
| SIF 2 certification assembly | SIL | 2 | 2 | SENS | 2338.3 | 4.28E-04 | 3174.94 | 1 | 3 | N/A | 4000 | | |
| | | | | LS | 24375.5 | 4.10E-05 | 1206.33 | | 3 | N/A | | | |
| | RRF | 100 | 141.9 | FE | 151.9 | 6.58E-02 | 187.16 | | 2 | N/A | | | |
| | | | | SIF | 141.9 | 7.04E-02 | 154.15 | | 2 | N/A | | | |

**DOES NOT met SIL2**

Option 1: Redundant valve
Option 2: PTI 4/year

FIG. 6. SIF analysis from proprietary software[b].

| | | Cost | Total SIF cost over plant life |
|---|---|---|---|
| **SIF 1 traditional assembly** | Total design and implementation | $6,300 | $597,790.97 |
| | Annual operation and maintenance | $12,656.27 | |
| | Net present value of annual | $373,664.13 | |
| | | Cost | Total SIF cost over plant life |
| **SIF 2 certification assembly** | Total design and implementation | $1,200 | $331,994.32 |
| | Annual operation and maintenance | $4,449.81 | |
| | Net present value of annual | $131,375.37 | |
| | | | Lifecycle savings $265,796.65 |

**FIG. 7.** Lifecycle cost comparison.

**Authors' company's certification scheme's impact on SIF analysis.** The goal of implementing a safety instrumented system (SIS) is to ensure that operations are safe, that equipment is available, and that the necessary risk reduction has been achieved. While this is true, it must be balanced with cost-effective solutions. Through SIF analysis, utilizing the authors' company's certification program assemblies reduces the PFDavg contribution and increases the mean time to fail spurious (MTTFS) of the final element—thus reducing false trips and significantly decreasing overall SIF lifecycle cost.

In this SIF analysis (**FIG. 6**), the same devices and assumptions have been maintained for both SIF 1 and SIF 2. The PFDavg contribution of the final element has been modeled for both SIF 1 (traditional assembly with individually certified devices) and SIF 2 (authors' company's certified assembly). In this analysis, SIF 1 achieves a Safety Integrity Level 1 (SIL 1) designation, which does not meet the target of SIL 2. SIF 2 (author's company's certification scheme) achieved a SIL 2 designation. Two options are considered to bring SIF 1 up to target:

- Option 1: Add an additional redundant final element assembly
- Option 2: Increase the proof test interval (PTI) to four times per year.

Both options add significant cost.

The authors' company's lifecycle cost comparison tool[b] (**FIG. 7**) demonstrates significant savings when utilizing the authors' company's certification's certified final element assemblies' SIF 2. In this example, a savings of 45% was achieved for an overall lifecycle cost reduction of $265,796.

**Takeaway.** While devices utilized in remote actuated valves typically have IEC 61508 certification, it has been recognized that more can still be done. The IEC 61508 concepts can (and should) be applied to the entire assembly. The authors' company's certification does this. It is a credible assessment and certification process that can verify, validate and document improved remote actuated valve assemblies. Perhaps the result will be that remote actuated valves will no longer win the first place of safety reduction. All end users will enjoy better products that will significantly improve safety, reduce false trips and decrease overall lifecycle costs. HP

**NOTES**

[a] exida's Remote Actuated Valve Assembly (eRAVA) certification
[b] exida's exSILentia integrated safety lifecycle tool

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# Double-suction centrifugal compressor in ethylene plant inlet piping design

Centrifugal compressors are designed with different suction nozzle orientations and impeller layouts in casing. Centrifugal compression trains may be supplied in series and parallel arrangements to achieve the total flow and head requirements of a specific application.

The impeller configuration and arrangement in centrifugal compressors can be straight-through, compound, back-to-back, double-suction, and side-loaded, among others.

Centrifugal compressors are used in ethylene plants for cracked-gas compression and refrigeration services. The large gas volume requirement in ethylene plants, and the need to reduce the head per stage due to fouling, make a double-flow configuration suitable for the first process stage of a cracked-gas compressor and for refrigeration compressors in ethylene plants.

This article will address the importance of suction piping design for double-suction centrifugal compressors in the compressor performance used in ethylene plants.

**Double-suction compressors.** A double-flow compressor is basically two smaller compressors arranged back-to-back with a common discharge (**FIG. 1**). The gas flow enters by two separate suction piping at two suction nozzles located in both ends of the compressor, is compressed by each impeller at each end, and then enters the double-flow wheel at the center of the compressor. This configuration enables a much higher volume flow within a smaller and less-expensive compressor.

The pressure rise across a double-flow centrifugal compressor is close to 2:1. The double-flow arrangement is located inside a single casing—normally, an auxiliary split. The gas exits from the double-flow compressor casing (low-pressure casing) and enters the second casing. The volume of the combined flow exiting the double-flow compressor is about equal to that of the stream entering each half of the double-flow compressor (double the mass flowrate at twice the pressure means that the volumetric flowrate is very similar). By this configuration, both the first and second casing can be driven by a single steam turbine.

The maximum number of stages per section is limited to four. It is possible to design the casing to a single section of compression, with no more than two suction nozzles and a single-discharge nozzle or two separate discharge wheels.

**Inlet piping to double-suction compressors.** Compressor performance and reliability depends on the suction piping design. The suction piping design should provide a uniform flow distribution to the first-stage impellers. Compressors are normally designed assuming a relatively uniform velocity profile at the compressor inlet flange. A careful and close collaboration between the compressor designer and the piping designer of the engineering company should be considered during the detail design of double-suction compressors. The straight-run piping requirement for axial inlet compressors is normally 10 times that of the inlet pipe diameters.

A double-suction compressor is like two common compressors in parallel to increase capacity of a system. In parallel centrifugal compressors with the same model, there is always a risk that both units will not be operating at the same point on the performance curve. This is because, even if they are "identical" units, they are not always the same, and they have different dimensions with different piping system resistance. Therefore, in parallel operation, it is always recommended that each compressor has a separate anti-surge system to protect each compressor. However, in a double-flow compressor, this is not easy or practical because the compressor has a common discharge nozzle. Thus, the design of the inlet piping must be performed carefully to achieve a well-balanced, stable and distortion-free flow into each suction nozzle of the compressor. If the piping design is not carefully performed, there is a high risk that the flowrates to each suction nozzle may not be balanced, thus causing problems such as premature surging. Two possible economic options exist for suction piping design for a double-flow compressor: collector design and Y configuration, which will be explained in the following sections.

**Collector design.** The most reliable inlet piping design for a double-flow compressor is by adding a collector or drum with two separate branches from the collector to split the flow (**FIG. 2**). This will cause well-balanced and distortion-free flow into each suction nozzle of the compressor casing. In **FIG. 2,** the collector with diameter D2 is added, the inlet pipe (D1) from the suction knockout (KO) drum is connected to the collector, and then two suction piping with diameter

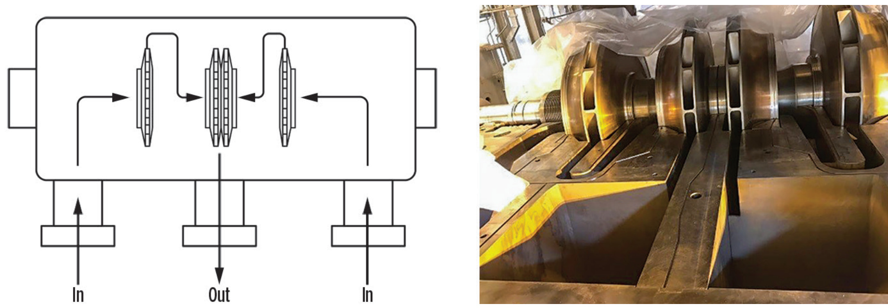


**FIG. 1.** Double-suction arrangement. Source: Google images.

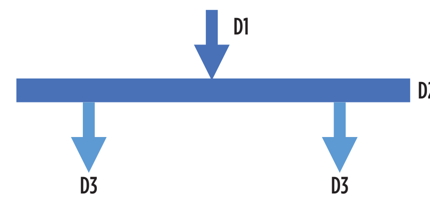


**FIG. 2.** Collector or drum arrangement.

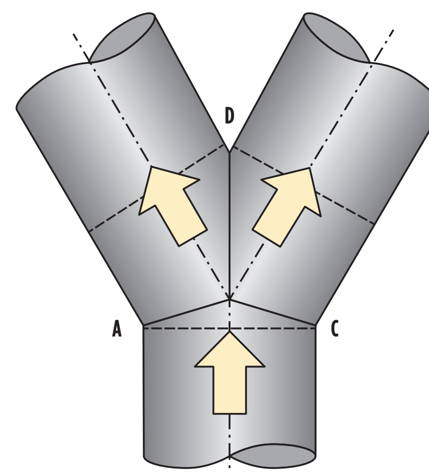


**FIG. 3.** Y-type splitter arrangement. Source: Google images.

D3 is branched from the collector and is routed to the compressor suction nozzles. In case D1 = D3, the risk of liquid buildup in each leg should be checked by a process engineer, and provision for collection of the liquid should be foreseen.

**Y configuration.** Another type of suction piping design is using a Y configuration (**FIG. 3**). This configuration can be fabricated in the field or in a vendor's shop. It is a type of miter design; however, it must have careful and precise dimensions, fit-up and welding.

This arrangement is most preferred by engineering companies due to lower installation costs. However, the use of miter is sometimes prohibited in the client piping material specifications or project piping branch tables. In addition, this arrangement is applicable for flanges with a maximum pressure rating of 300. Using a Y configuration may cause unequal flow distribution to the compressor inlet. If one leg of the Y has less flow vs. the other leg, then this may result in one section of the compressor running near surge, while the other section runs and operates near the overload region.

When using this arrangement, the following is recommended:

- A proper upstream straight run of pipe of 10D–12D should be considered between the neck of the Y connection at the inlet and the weld point of the upstream elbow. Otherwise, any disturbance in the piping upstream of the Y will cause the flow to shift to one leg of the Y or the other side.
- Try to fabricate the Y in a controlled shop with a high-quality weld and dimensional inspection. Any discontinuity (e.g., weld penetration or pipe penetration) in the internal surface of the Y—especially at points A and C (**FIG. 3**)—may cause flow separation from the wall and shift the gas flow toward the other leg. The use of internal coating in the suction piping between the suction KO drum and compressor is recommended to be studied and decided.
- Piping legs from the drum/collector or Y assembly to each compressor inlet nozzle must be identical to each other, with identical slope, elevation and length. Each leg should create the same resistance to the gas flow to minimize the risk of gas switching to the low-resistance leg/path.
- For a Y-type splitter assembly, the large radius at the dividing point/angle B is preferred. A mitered type of Y joint with a sharp, pointed dividing geometry could cause flow separation and uneven distribution of gas between the legs.
- Low gas velocity will ensure equal flow distribution.
- Install the Y assembly close to the compressor casing and consider a straight pipe run downstream of the Y assembly or collector, based on the compressor manufacturer's recommendation. In this case, it is recommended to ask the compressor manufacturer to review the inlet piping.

**Other considerations.** Based on specific compressor requirements, it is recommended to study the following requirements:

- The strainer location close to the Y splitter assembly or collector may cause flow separation. A T-type strainer is recommended here. It is not recommended to use a cone-type strainer for suction sizes above 12 in. to minimize the cost and risk of leakage from extra flanges. The strainer should be properly designed and braced to minimize the risk of collapse and to mitigate clogging.
- Check with the vendor to see if the location of the compressor's surge control sensors should be on both legs close to the compressor or upstream of the Y splitter assembly or collector. Being upstream of the Y splitter assembly or collector may cause an error, as there is the possibility that one leg could be overloaded, and that the other leg could have less flow.
- Do not install instruments (such as temperature gauges or temperature transmitters) downstream of the strainer. This is a design mistake. It puts the compressor at risk if a thermowell is broken due to the high-frequency acoustic vibration and from carryover of gas into the compressor.
- Consult with the compressor manufacturer on the proper

**FIG. 4.** Y-type splitter arrangement for a propylene gas compressor.

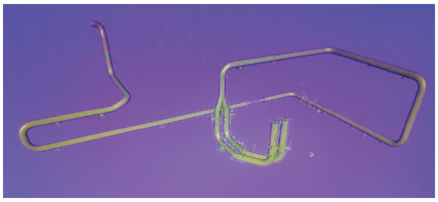

**FIG. 5.** Y-type splitter arrangement for a double-suction compressor.

location of the strainer and of the valve at the suction piping. Using an individual valve for each leg is not common practice and will increase costs. However, to isolate the compressor immediately after a seal failure, the proper number and location of the isolating valves should be studied and confirmed by the vendor.

- The minimum instruments at the suction pipe should be a temperature gauge/transmitter, a pressure gauge/transmitter, and a pressure differential gauge and transmitter across the strainer. In addition, a pressure differential transmitter should be included to measure the pressure at the impeller inlet to the pipe upstream of the compressor flange.
- The need for level transmitters at each leg close to the compressor should be studied. If necessary, add a trap for the strainer bottom.
- It may be required to add some type of trimming device (such as an orifice plate or a butterfly valve) in one or both legs of the double-flow compressors to equalize flow. The most economical method may be to install a valve upstream of a Y splitter to control flow.

**Case study.** During commissioning, a double-flow propylene compressor (**FIG. 4**) was found to have icing on one suction nozzle and no icing on the other prior to compressor startup. After compressor startup, the compressor had low head pressure and it surged prematurely. The investigation showed that the inlet piping caused unequal flow distribution to the compressor inlet. This resulted in one section running near surge, while the other section was operating near the overload region. The inlet piping to the compressor was modified to have the same and symmetrical dimension and elevation. In addition, the Y splitter (**FIG. 5**) was installed closer to the compressor and was modified to have symmetrical legs and a larger radius at the dividing point B leg angle (**FIG. 3**), as well as an internal coating and identical dimension. All temperature instruments were relocated upstream of the strainer.

**Takeaway.** Experience and knowledge are imperative when designing the inlet piping of double-suction compressors. A Y-splitter design should not be considered as a piping miter, which can be manufactured onsite during construction. It is recommended to ask the compressor vendor to supply and design the Y splitter and the associated downstream piping, or to at least review the associated piping design drawings. The need for an internal coating of the inlet piping to the compressor should also be studied. HP

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B. FORBES, AMETEK, Berwyn, Pennsylvania

# Strategies to minimize piping thermal maintenance system cost without compromising performance

Where process fluid must be maintained within a certain operating temperature window, the piping will likely require the application of a thermal maintenance system. Thermal maintenance systems can utilize a broad range of technologies—from bare tube tracing to fully jacketed pipe. However, regardless of the technology employed, engineering must be performed to design the system. The approach used for the engineering and design of the system can have a dramatic impact on the cost. This article presents four strategies for minimizing the total cost without compromising performance. The strategies are:

1. Match the heating technology to the application
2. Optimize the heating circuit lengths
3. Optimize the utility infrastructure
4. Structure the bid process to reward optimization.

In addition, three real-life examples are presented that demonstrate the actual savings that these strategies can achieve. These examples are:

- Example A showing the benefit of Strategies 1 and 2 combined
- Example B showing the benefit of Strategy 3
- Example C showing the benefit of Strategy 4.

While much of this discussion centers around steam heating systems, the same principles apply to other heating fluids and cooling systems.

**Strategy 1: Match the heating technology to the application.** Different processes require different considerations in the design of the thermal maintenance system. The most common considerations are:

- **The purpose of the heating system:** Liquids must often be maintained above a freezing point. However, for vapors, condensation within the piping is often a primary concern. In some applications, a pre-heat or recovery (melt-out) condition is the primary concern.
- **The acceptable temperature window:** Some processes must simply be maintained above or below a critical threshold. Other processes have both an upper and lower bound that must be considered.
- **The available temperature differential:** Some applications afford a large "temperature delta" that enables even a relatively poorly performing technology to be used successfully [e.g., maintaining a pipe above the water freezing point (0°C) using 10 barg steam (184°C)]. Other applications provide very little "temperature delta" and require a very robust heating system to ensure success [e.g., maintaining a pipe at 180°C using the same (184°C) steam].

Properly matching the heating technology to the application is key to reducing cost. An over-performing heating technology will cost more than a performance-matched technology. Conversely, applying an under-performing technology will necessitate the use of additional tracing runs and heating circuits, which will drive up the cost of the utility components. A properly matching technology will optimize both costs: (1) the heating technology itself will cost no more than what is necessary to achieve the purpose and (2) the utility infrastructure requirements will be minimized along with associated costs.

Many tracing technologies are available on the market. A true technology partner can help you select the best option without bias to any one solution. FIG. 1 is one example of a robust range of heating technologies.

Technology 1[a] provides an eight times

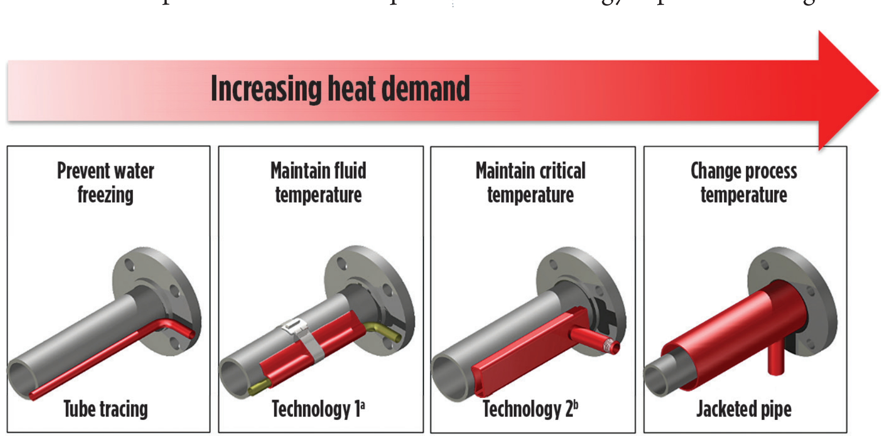


FIG. 1. One example of a heating technology suite.

increase in heat transfer over tube tracing. Technology 2[b] provides a two times increase over Technology 1, as well as more consistent performance for critical applications. Jacketed pipe has no contact resistance; therefore, heat transfer is limited primarily by the convection coefficients of the fluids.

**Strategy 2: Optimize the heating circuit lengths.** Properly matching the heating technology to the application is only the first step. To keep costs as low as possible, the heating technology must be used to its fullest potential. Correctly sizing the system—in terms of the number of tracer runs per pipe length—is the most obvious aspect of this. However, it is equally important to maximize the length of heated piping between each supply and return point. Each of these heating fluid flow paths are commonly referred to as circuits (FIG. 2).

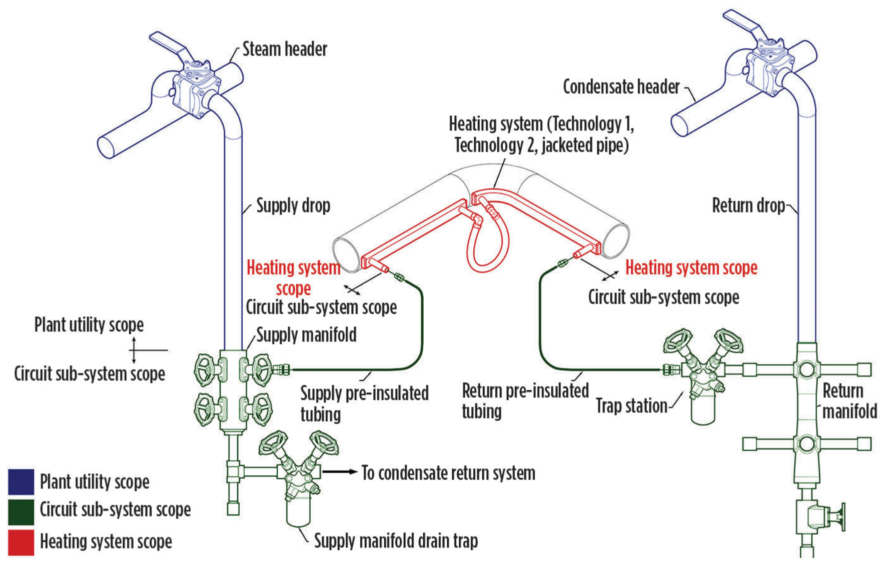


FIG. 2. Components of a steam tracing circuit.

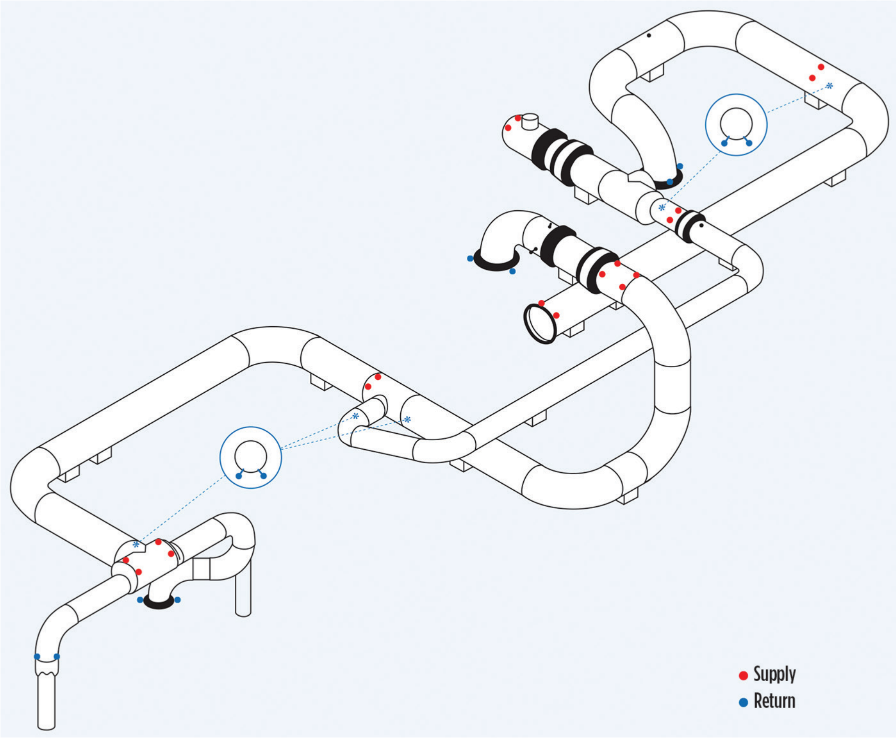


FIG. 3. Example piping system with circuit supply and return points.

The heating fluid loses temperature as it travels through the circuit. In the case of steam, the temperature loss is driven by pressure drop. If a circuit is too long, the heating fluid temperature towards the end will be too low to accomplish the thermal objective. One approach to avoid this issue is to conservatively limit the length of all the circuits. This approach is commonly the basis of plant tracing specification documents and is also often employed by tracing vendors to avoid the burden of additional engineering. This approach works because there are no technical concerns with having a circuit that is "too short." However, it does require an unnecessarily high number of heating circuits, which then requires additional utility infrastructure, along with the associated costs.

A preferred approach is to establish each circuit length based on an actual calculated temperature drop. At a minimum, this calculation must be performed for each unique combination of process condition and lines size. Performing this calculation will create additional engineering cost, but that cost is more than offset by the savings achieved by using a reduced utility infrastructure. In many cases, plant tracing specifications are so conservative that using calculated circuit lengths can reduce the number of circuits fourfold.

**Strategy 3: Optimize the utility infrastructure.** Any heating system applied to piping requires periodic supply and return points. The location of these points is governed by piping geometry and the circuit length limitations. The resulting system will have supply and return points scattered throughout the facility. Proximate supply and return points are typically grouped together and connected to common supply/return manifolds (FIG. 3).

Each manifold must be mounted at an accessible location and plumbed into the plant's supply/return headers. The total cost of each manifold is significant, which is why it is desirable to keep the total number to only what is required. However, accomplishing this practically is a challenge. The hundreds of supply/return points and thousands of possible manifold locations form a jigsaw puzzle with millions of possible solutions.

The traditional approach is to scatter manifolds throughout the plant based

on rough approximations and past experiences. The heating system supply/return points are then tied in wherever they can be. Where they cannot be tied in, last-minute manifold additions become necessary. Similarly, where manifolds or ports are not needed, the paid-for hardware sits unused.

A better approach is to match the manifolds to the circuits during placement. The goal of this approach is to minimize the number of manifolds by maximizing the number of circuits connected to each manifold. For example, the author's company has developed a proprietary algorithm[c] that analyzes the supply and return points in 3D space and determines the optimum manifold locations that result in the smallest total number of manifolds.

This tool provides customers full flexibility to restrict the analysis to only acceptable manifold locations, to restrict the supply and return tubing length and to include existing manifolds in the analysis. With this tool, the resulting system design will always be the most optimal manifold placement solution.

The manifolds shown in FIG. 4 are the most common in steam heating systems. They are less common when the heating fluid is a liquid. Nevertheless, many of the same concepts of optimization can apply regardless of the specific supply and return hardware used.

**Strategy 4: Structure the bid process to reward optimization.** The most powerful tool that can be used to control costs and optimize system design is to structure the contracts so these goals are rewarded. For example, in the case of a pipe heating system, the contract should reward the application of the three optimization strategies previously described. Two commercial mechanisms should be considered to accomplish this:

1. **Requires lump-sum bids—** When a vendor is contracted to supply components at a per-unit rate, the vendor is not motivated to reduce the provided hardware quantity. Instead, vendors should be required to provide lump-sum bids that clearly define deliverables and performance guarantees. This motivates the vendor to minimize the quantities of hardware provided to maximize their profit. This approach will only be successful if the bid includes a commitment to well-defined deliverables and performance criteria. For pipe heating systems and associated hardware, consider requiring the vendor to supply the following:
   - A clear statement of thermal objective for each line (for performance accountability)
   - Calculations/thermal models justifying the quantity of tracers applied
   - Calculations/thermal models justifying the tracer circuit lengths specified
   - Detailed installation drawings for the tracing system and other supplied hardware
   - Field support for and inspection of the installation
   - A performance guarantee with defined corrective action
   - Material specification and fabrication standards for all supplied hardware.
2. **Consolidate scope to a small number of vendors—**A vendor can only optimize what is within their scope. They do not necessarily have good insight into how decisions about their scope affect others. One vendor's choice to optimize their scope may necessitate a disproportionate increase in another. A common example is tracing circuit lengths. The tracing vendor may be able to reduce their cost by using fewer tracer runs on the pipe by using very short circuit lengths. This can easily drive up the number of circuits, resulting in a disproportionate increase in the cost of the utility infrastructure. The solution is to consolidate the scope of supply to a smaller number of vendors. When one entity is responsible for the total scope, optimization decisions are

The most powerful tool that can be used to control costs and optimize system design is to structure the contracts so these goals are rewarded.

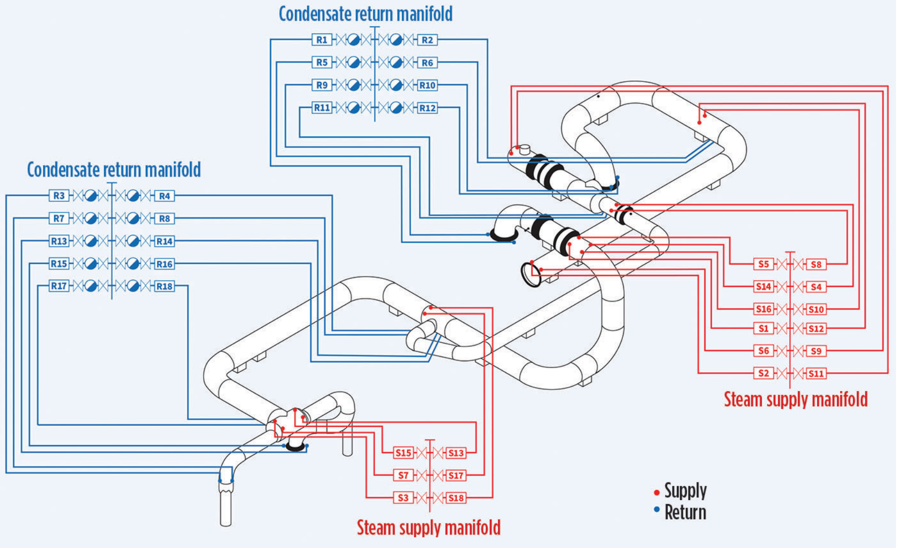


FIG. 4. Example piping system with one possible manifold grouping indicated.

made with consideration for the complete system. Like the first strategy, this approach is most successful when coupled with accountability to provide well-defined deliverables.

**Cost savings.** The savings achieved by these four strategies can be broadly categorized into three buckets:

1. **Savings achieved by reducing the quantity of hardware.** Applying the engineered approach instead of conservative specifications will significantly reduce the number of parts required. This reduction is far more significant than any bulk buy or discount pricing effect could achieve.
2. **Savings achieved by reducing the number of last-minute purchases and project delays.** The impact of this is much more difficult to quantify, but experience shows it to be significant. A project that is well planned prior to execution experiences fewer delays and surprises.
3. **Savings achieved by reducing the long-term operating costs.** Long-term operating costs are primarily associated with energy utilization and parts replacement. Reducing the number of parts that comprise the heating system has a significant effect on both.

**TABLE 1.** Applying recommendations resulted in an 89% reduction in the number of steam circuits

| | **Original design basis** (Bare tube tracing only) | **Revised design** (Combination of bare tube tracing and Technology 2[b]) |
|---|---|---|
| Total length of pipe, ft | 7,853 | 7,853 |
| Total length of tracing, ft | 96,351 | 11,482 |
| Total number of circuits | 2,600 | 274 |

**TABLE 2.** Technology 2[b] provided an overall cost savings of 78%

| | **Original design basis** | **Revised design** |
|---|---|---|
| Cost of the heating system | $1,342,000 | $2,073,000 |
| Cost of installing the heating system | $589,000 | $206,000 |
| Cost of utility infrastructure | $16,105,000 | $1,973,000 |
| Cost of utility install | $7,731,000 | $1,223,000 |
| **Total cost** | **$25,193,000** | **$5,475,000** |

**TABLE 3.** Additional utility infrastructure also creates additional operating expenses; these come in the form of additional steam consumption and maintenance costs to maintain the traps

| | **Original design basis** | **Revised design** |
|---|---|---|
| Heating system steam load, lb/hr | 8,600 | 8,600 |
| Utility infrastructure steam load, lb/hr | 53,500 | 5,200 |
| Total steam load, lb/hr | 62,100 | 13,800 |
| **Yearly steam cost** | **$4,352,000** | **$967,000** |
| Number of circuits | 2,600 | 274 |
| Trap replacements/yr | 433 | 46 |
| **Yearly trap cost** | **$507,000** | **$53,000** |

**TABLE 4.** Cost basis for the Canadian chemical project

| **Item** | **Cost** |
|---|---|
| 12-port high-pressure supply manifold | $7,270.10/each |
| 0.5-in. tubing x NPT adapter | $15.65/each |
| 0.5-in. tubing x JIC adapter | $37.05/each |
| Tubing end seal kit | $52.95/each |
| 0.5-in. stainless pre-insulated tubing | $7.18/ft |
| 0.5-in. tubing x tubing union | $34.65/each |
| 12-port high-pressure return manifold | $28,619.46/each |
| 0.5-in. stainless tracer tubing | $1.75/ft |
| Labor rate | $50/hr |
| Steam | $8/1,000 lb |
| Replacement of high-pressure steam trap | $1,120/each |
| Steam trap service life | 6 yr |

**Example A: Optimizing technology selection and circuit lengths (Strategies 1 and 2).** The author's company provided a heating system for a large, newly constructed, Canadian chemical plant. The plant design was like a previous plant that exclusively used bare tube tracing. It was initially expected that the same would be used in the new plant. The author's company reviewed the basis of the technology selection and recommended two cost-saving changes:

1. Use proprietary stream tracing Technology 2[b] instead of tube tracing on lines with more stringent thermal requirements (i.e., lines requiring many tube tracing runs)
2. Establish circuit lengths based on calculated steam pressure drop instead of the previous plant specification.

Applying these recommendations resulted in an 89% reduction in the number of steam circuits (**TABLE 1**). Switching a portion of the project to proprietary Technology 2[b] increased the cost of the heating system but significantly reduced the number of circuits, providing an overall cost savings of 78% (**TABLE 2**).

The utility infrastructure includes the following components:

- Steam supply manifolds and condensate return manifolds
- Supply and return tubing for each

circuit (tubing connecting steam supply/return manifolds to circuit supply/return points)
- Steam trap for each circuit
- Steam trap for each supply manifold
- Fittings and hardware required to connect all the above components.

**Note:** In this example, proprietary Technology 1[a] tube tracing could have also been used as an intermediate technology between the tube tracing and the proprietary Technology 2[b]. Doing so would have resulted in additional cost savings. However, the customer chose not to pursue this option as they preferred the simplicity of using a smaller number of technologies. Additional utility infrastructure also creates additional operating expenses. These expenses take the form of additional steam consumption to support the extra heat load, and the form of additional maintenance cost to maintain the traps (**TABLE 3**).

The cost basis of this project's individual items is summarized in **TABLE 4.** Note that this project utilized high-pressure steam, which required more expensive manifolds and steam traps. A more typical steam pressure would have resulted in lower utility infrastructure costs and lower operating costs for both options. The impact of this change on purchase cost would be relatively small, with savings shifting from 78% to roughly 76%. The impact on operating cost would be more significant, with savings shifting from \$3.4 MM/yr to \$2.1 MM/yr for steam cost and \$454,000/yr to \$175,000/yr for trap maintenance.

This and the following examples all use pre-insulated tubing to supply and return the heating fluid to the tracing. Pre-insulated tubing typically has a higher purchase cost than traditional hard pipe, but it provides significant time savings during installation.

**Example B: Optimizing utility infrastructure (Strategy 3).** The author's company provided a heating system for a major Middle Eastern refinery expansion project. The engineering, procurement and construction contractor had already placed manifolds throughout the plant before the author's company was engaged on the project. The author's company was asked to optimize the manifold utilization and applied its proprietary algorithm[c]. The expectation was that some portion of the planned manifolds would prove unnecessary and could be removed from the plan. This expectation proved to be true; optimization resulted in the removal of 15% of the manifolds and a total cost reduction of 43% (**TABLE 5**).

While the cost savings are significant, employing the proprietary algorithm on most projects will yield even more significant savings. Two factors affecting this project that limited the cost savings potential were:

1. The planned manifold locations were based on a previous project for a very similar plant utilizing the same heating technology. Thus, the planned locations were actually very good approximations of what would be needed, much better than on an average project.
2. Only manifold locations that were already planned could be used. The author's company was engaged late in the project cycle. Planned manifolds could be eliminated, but adding manifold locations was not an option. Thus, the resulting manifold locations were not optimized to the extent that they could have been.

The cost basis of this project's individual items is summarized in **TABLE 6.** 0.75-in. tubing is used for each circuit supply run and 0.5-in. tubing is used for each circuit return run.

**Example C: Optimizing the bid structure (Strategy 4).** The author's company provided a bid for another major Middle Eastern refinery expansion project. This bid included the application of all three optimization strategies:

1. Optimization of technology by using a combination of tube tracing and proprietary Technology 1[a] in place of tube tracing alone.
2. Optimization of circuit lengths via engineering calculations.
3. Optimization of manifold quantity and placement via a proprietary algorithm[c].

The lump-sum bid was being compared to another vendor's per-unit bid. The

**TABLE 5.** Optimization techniques provided a cost reduction of 43%

| | Planned | Optimized |
|---|---|---|
| Number of manifolds | 186 | 159 |
| Number of used ports | 1,116 | 762 |
| Total length of tubing, ft | 225,000 | 80,000 |
| **Total cost** | **\$3,187,000** | **\$1,820,000** |

**TABLE 6.** Cost basis for a heating system for a Middle Eastern refinery expansion project

| Item | Cost |
|---|---|
| 4-port supply manifold | \$2,540.82/each |
| 8-port supply manifold | \$3,787.26/each |
| 12-port supply manifold | \$5,250.84 |
| 4-port return manifold | \$948.93/each |
| 8-port return manifold | \$1,731.48/each |
| 12-port return manifold | \$2,499.93/each |
| Trap station with steam trap | \$964.44/each |
| 0.75-in. stainless-steel, pre-insulated tubing | \$6.56/ft |
| 0.5-in. stainless-steel, pre-insulated tubing | \$4.69/ft |
| 0.75-in. tubing x NPT adapter | \$32.67/each |
| 0.75-in. tubing union | \$55.22/each |
| 0.75-in. tubing x JIC adapter | \$69.67/each |
| 0.5-in. tubing x 0.75-in. NPT adapter | \$27.36/each |
| 0.5-in. tubing union | \$33.97/each |
| 0.5-in. tubing x 0.75-in. JIC adapter | \$55.78/each |

end user chose to use the other vendor due to the lower apparent per-unit cost and the relatively late engagement with the author's company. After the project was completed, the author's company was provided a tabulation of the actual hardware used on the project. Unsurprisingly, although the project scope had not increased, the number of components used was considerably higher than what was initially estimated. The final cost of the as-built system far exceeded the author company's lump-sum proposal (TABLE 7).

**TABLE 7.** The as-built system vs. the author company's lump-sum bid

| | **As-built system** (Unit pricing) | **Author company's proposed system** (Lump sum) |
|---|---|---|
| Engineering | – | $35,010 |
| Tracing system | $1,075,801 | $786,814 |
| Manifolds, traps and valves | $7,138,942 | $2,925,358 |
| S/R tubing and adapters | $3,399,362 | $1,812,481 |
| Installation labor | $6,370,765 | $3,504,970 |
| **Total cost** | **$17,984,870** | **$9,064,633** |

The per-unit pricing created a motivation for the vendor to allow the project scope to balloon. A lump-sum bid would have ensured that costs increase only in response to a legitimate expansion of scope. By using the lump-sum approach, the end user could have saved 50% on the project.

The cost basis of this project's individual items is summarized in TABLE 8. The as-built system utilized a combination of 0.375-in., 0.5-in. and 0.75-in. tube tracing. The optimized system utilized a combination of 0.5-in. tube tracing and the proprietary Technology 1[b].

**TABLE 8.** Cost basis of a major Middle East refinery expansion project (Example C)

| **Description** | **Cost** |
|---|---|
| 0.375-in. stainless-steel tracer tubing | $2.76/ft |
| 0.5-in. stainless-steel tracer tubing | $3.34/ft |
| 0.75-in. stainless-steel tracer tubing | $6.65/ft |
| Technology 1[a] enhancer | $7.19/ft |
| 0.375-in. tubing union | $13.71/ft |
| 0.375-in. tubing x 0.75-in. SW adapter | $19.42/each |
| 0.375 in. x 0.5 in. tubing union | $21.04/each |
| 0.5-in. tubing union | $20.42/each |
| 0.5-in. tubing x 0.75-in. SW adapter | $21.99/each |
| 0.5-in. tubing x NPT adapter | $27.89/each |
| 0.75-in. tubing union | $32.99/each |
| 0.75-in. tubing x SW adapter | $24.93/each |
| 0.75-in. nipple | $9.24/each |
| Tracer banding | $0.49/ft |
| Tracer banding buckles | $0.42/each |
| Heat tracer compound | $61.60/gal |
| 4-port return manifold | $1,627.78/each |
| 8-port return manifold | $2,220.68/each |
| 12-port return manifold | $2,887.50/each |
| 4-port supply manifold | $2,962.96/each |
| 8-port supply manifold | $4,288.90/each |
| 12-port supply manifold | $5,824.28/each |
| 0.5-in. plug | $8.78/each |
| 0.375-in. stainless-steel, pre-insulated tubing | $4.90/ft |
| 0.5-in. stainless-steel, pre-insulated tubing | $4.69/ft |
| 0.75-in. stainless-steel, pre-insulated tubing | $6.56/ft |
| Pre-insulated tubing end seal | $7.64/each |
| 0.75-in. circuit isolation valve | $115.50/each |
| Steam trap | $934.78/each |

**Takeaways.** Controlling the cost of a thermal maintenance system requires the vendors to apply an engineered approach. The key focus of this engineering approach is to:

1. Match the heating technology to the application
2. Optimize the heating circuit lengths
3. Optimize the utility infrastructure.

To ensure that the vendor(s) applies this approach, the bid should be structured in such a way that optimization is rewarded. This is accomplished by:

1. Requiring lump-sum bids with well-defined deliverables and performance criteria
2. Consolidating the scope to a smaller number of vendors to achieve cross-discipline optimization.

To best capitalize on these strategies, look for a technology-neutral thermal maintenance system provider with strong engineering capabilities. HP

**NOTES**

[a] Controls Southeast Inc.'s (CSI's) TraceBoost technology
[b] CSI's ControTrace technology
[c] Manifold Optimization Scheme

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# Down with downtime: Repair complex petrochemical components with selective electroplating

The volatile impact of the coronavirus pandemic hit at the same time the petrochemical industry was entering a downcycle. Demand has already started to slow, while a surplus in capacity is eroding margins and value. The global disruption that the pandemic has brought will only serve to further exacerbate these issues.

Several factors are at play, but the key one is the global clampdown in the use of plastics. The damage plastics cause to the environment and oceans is well documented, and more countries and businesses are cutting back on these petrochemically produced products. Indeed, 250 organizations—collectively responsible for 20% of the plastic packaging produced around the world—have committed to reduce waste and pollution under the New Plastics Economy Global Commitment.[1]

Meanwhile, billions of dollars' worth of refineries and processing plants have come, or are about to come, online. Between 2014 and 2019, $120 B was spent on new facilities in Texas, much of it focused on producing high-density polyethylene.[2] At least 108 petrochemical plants are scheduled to start operations in Asia and the Middle East by 2023. With the cutback in plastics production, this is causing a perfect storm of global overcapacity, and analysts predict the bust will happen in as little as five years.

However, not all petrochemicals will be affected. According to Deloitte's *The Future of Petrochemicals* report, certain petrochemicals—such as ethylene and propylene—are still predicted for growth and have a high utilization rate. However, shutdowns and maintenance activities may inhibit current capacity required to sustain the production needed to keep up with demand, thereby introducing volatility in the price of base chemicals.

**Minimizing downtime is imperative.** Minimizing downtime is vital. Operators cannot afford for plants to be down longer than their scheduled maintenance periods—or worse, be plunged into unplanned downtime—as this would result in huge losses in productivity and profits.

Maintenance planners and turnaround managers must be able to control and plan maintenance work, and the ability to estimate time and resources is essential. Methods of repair must also be quality assured, reliable and cost effective.

A wide range of methods are available, each with varying benefits and challenges. When it comes to components with complex geometries that must be repaired fast and must be able to withstand the harsh, continuous operating conditions of petrochemical production, few solutions are as adept and as effective as selective electroplating.

**Selective plating: Serving many functions, suitable for many applications.** Selective electroplating, such as the author's company's proprietary selective electroplating process[a] is a proven, efficient and economical way of performing surface treatment repairs. The proprietary process is a portable plating method used to enhance, repair and refurbish localized areas on manufactured components.

Plating can serve a variety of purposes, such as a localized defect repair or bringing an inside diameter (ID) or outside diameter (OD) back to size. Plating can also enhance wear or corrosion resistance exactly where it is needed, even on new parts where it would be prohibitive to tank plate the entire part.

The process uses fundamental electrochemical principles. An electrolyte solution, which contains ions of the metal to be deposited, is introduced between the negatively charged part to be plated and the positively charged plating tool (i.e., anode). A portable powerpack provides the required direct current and allows precise control of amperage, voltage and plating time for high-quality and accurate plating results.

The circuit is completed when the anode touches the surface of the part to be plated. A suitable cover material around the tool provides a reservoir to evenly distribute the electrolyte. The current causes the metal ions in the electrolyte to bond with the surface of the part and build up the plating layer. The result is a highly adherent and dense metal deposit. The metal or alloy to be deposited can be chosen from more than 50 different solutions, which allows the application to be tailored to the desired characteristics of the plating material.

The following are a few examples of where selective electroplating can commonly be used effectively in the petrochemical industry:

- Corrosion protection
- Defect repairs
- Dimensional restoration
- Wear resistance
- Improved hardness
- Prebraze
- Anti-galling and slip.

**Quality of selective plating.** One of the key criteria that maintenance and turnaround managers need to consider is quality. Here, selective plating gives the assurance needed due to the deep, strong molecular bonds it creates between the coating and the substrate. Along with this benefit, repairs conform to rigorous quality standards and are performed by trained, highly skilled technicians.

**Judge by adhesion.** When judging surface coatings, a telling sign of its quality is its adhesiveness (i.e., how strong the deposit's bond is with the base metal). Typically, when comparing selective plating, there are two alternative surface coating choices: thermal spray coatings and metal-filled epoxy coatings. Against both, selective plating is superior in bond quality and performance.

Unlike selective plating, thermal spray coatings can only achieve a mechanical bond. This means the coating is not fused to the base metal, and adhesion comes from mechanical interlocking formed chiefly from the roughness of the substrate surface, making it comparatively weaker and much more susceptible to wear and corrosion.

**FIG. 1.** A proprietary coating[b] was applied to an OEM's valve assemblies.

**FIG. 2.** The proprietary coating[b] on the valve seat.

Conversely, metal-filled epoxy coatings are coatings that stick to the substrate like an epoxy glue. Generally, these coatings can only achieve an adhesion between 1,500 psi and 5,000 psi. In an adhesion test against selective electroplated deposits, the epoxy coating fails long before the electroplated coating delaminates. This is an indication that the adhesion of selective electroplated coatings is in excess of 10,000 psi.

Such strength and quality are achieved because of the deep molecular bonds that selective plating creates between the deposit and substrate. The plating nucleates on the existing metal surface, extending the crystal structure of the base metal.

**Quality standards.** Another mark of quality and performance is the standards and specifications the repair method has achieved. Selective plating can be looked at from a host of different industry specifications, including military standards, aerospace material standards and federal standards.

The proprietary selective electroplating solution has been used for more than 50 yr in oil and gas, industrial and military industries and adheres to numerous standards and specifications, including the AMS 2451 and MIL-STD-865_D standards.

It is also important to use trained technicians to perform the selective electroplating repairs. Technicians can control the plating process and, depending on the deposit solution, the process provides desired features such as hardness and corrosion, ensuring repairs are tailored to each unique situation.

**Reliability repairs to run to the next scheduled downtime period.** In tandem with quality and performance is reliability. With shutdowns scheduled 3 yr–6 yr in advance and in an industry that must run like clockwork, reliability is mission critical.

However, with petrochemical plants and machinery subject to steam and corrosion, along with wear and scoring, the question is how the components can be effectively protected to last from one downtime period to the next. Some common applications where repairs are often required include compressors, ball valves, gate valves, and casings and housings for pumps and turbines.

For scored, heavily worn or mis-machined components, copper can be electroplated onto the base metal to build the part back up to its original dimensions and tolerances. Meanwhile, a variety of nickel deposits can be used to provide corrosion protection, whether that is from chemicals or steam.

In many cases, restoring or enhancing components with complex geometries are not an issue, so long as the anode is able to touch the surface that is being plated. Repairs can be made to IDs, ODs, flat surfaces and more.

All repairs are evaluated on a case-by-case basis, considering disassembly, downtime and throughput, among others, to ensure the best possible outcome.

**Protecting 1,500 valves per year from steam and acid corrosion.** One such example of where reliability was increased due to selective electroplating was when the author's company helped with the plating of 1,500 original equipment manufacturer (OEM) valves per year. This was to prevent various valve assemblies from steam and acidic water corrosion.

Steam control systems are an essential part of refineries and large petrochemical plants. Corrosion resistant coatings are vital to valves and valve assemblies as the steam and acidic water found in these applications corrode the IDs of the valves.

However, plating these components comes with its challenges. In this application, the bore sizes of the valves varied in diameter from 1 in.–10 in., and turnaround time needed to be fast. This meant the alternative surface coating option (electroless nickel) was unsuitable.

Selective plating was chosen for this application because it was able to control the parameters and the deposit. The corrosion protection on the valves' IDs was much greater than what electroless nickel could achieve. The turnaround time was also much shorter due to less processing time, providing a better overall value.

Valves were prepared by first using general masking around the location of the ID to be plated. Using a proprietary coating[b] specially formulated for high corrosion resistance, the deposit was applied using ID

13's anodes and a special tool plating to a thickness of 0.001 in (**FIGS. 1** and **2**).

**Lowering costs through selective plating.** Selective electroplating is a cost-effective repair method for several reasons.

The first is that it is portable. Whether the job is in a shop or out in the field, selective plating can be conducted almost anywhere. It can also be mechanized or automated for minimal operator intervention. Repairs can be made in-situ for unplanned downtime situations, which mitigates the costs of disassembly, transportation and re-assembly.

Secondly, costs and downtime are also reduced due to the minimal amount of pre- and post-machining required. The specific area needing protection or repair must be masked off. By following the appropriate preparatory steps for the procedure, the deposit will adhere without issue. Due to the precision and accuracy of selective plating, there is often no need for post-machining, as the deposit can be plated to size.

Finally, repairs are fast. Once the repair case has been evaluated, and if machinery does not need disassembling, turnaround time can be as quick as a single working day.

**Leaving your sustainability credentials intact.** With increasing pressures across all industries to set and meet sustainability targets and reduce emissions, there is a growing focus on the environmental impact of unexpected shutdowns at petrochemical plants and refineries.

Plant interruptions can be highly damaging to the environment, and years' worth of gases and emissions can be leaked into the atmosphere from just a few hours of unplanned downtime. For example, gas flaring is one of the common side effects of unplanned shutdowns. Data published by the World Bank's Global Gas Flaring Reduction (GGFR) program shows that 145 $Bm^3y$ of gas is released into the atmosphere from gas flaring, equating to 270 MMtpy of $CO_2$ emissions. Therefore, getting facilities back online as fast as possible is crucial to lessening the environmental impact, and selective electroplating helps make this possible.

The proprietary electroplating process is highly sustainable, as well. In comparison to other surface coating methods (e.g., tank plating), selective plating uses much less solution and chemicals and generates very little wastewater. For workers and the environment, it is also much safer, with fewer fumes emitted and less hazardous waste to dispose.

With the portable nature of selective plating, there often is no need to ship components off for repair, which incurs transportation costs, more emissions and a bigger carbon footprint. Instead, technicians can bring portable selective electroplating equipment—containing all the solutions and accessories required to make the repairs—directly onsite.

**Managing maintenance in a new era.** The turn of the decade has brought many unforeseen events; however, the petrochemical industry was looking at a precarious new era long before the coronavirus pandemic arrived.

With a shrinking value pool, margin erosion and the changing way society views plastics, plants and machinery must be managed more carefully. Maintenance schedules are just one factor that petrochemical executives will need to be mindful of, but they are possibly the most important. However, by considering selective plating, maintenance planners can approach these situations with confidence. Due to each selective plating application being unique, repairs will be thoroughly considered and performed by trained, knowledgeable technicians, ensuring quality and reliability for years to come. HP

**NOTE**

[a] SIFCO Process®

[b] SIFCO ASC's AeroNikl® 250 coating

**LITERATURE CITED**


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# High-frequency acoustic excitation—Part 2

Failure of a piping section at the outlet of pressure safety valves (PSVs) and control valves—where significant differential pressure exists (particularly in gas processing industries)—is a common problem in industries. Most failures are caused by acoustic-induced vibration. High-pressure discontinuity generates high vibration and sound levels, causing the failure of the piping system. Generally, the likelihood of failure (LOF) is established through the calculation of sound power level (represented as PWL) at the source and at different sections of the downstream piping. Normally, it is the welded joint that fails due to acoustic excitation. The welded joint can be of any type, including a branch connection, welded tee, welded support, etc.

Basic understandings of high-frequency acoustic excitation have been explained in Part 1 of this work (December 2020). General calculation procedures for design sound power level ($PWL_D$) and sound power level at downstream discontinuity ($PWL_{dis}$) have also been explored. Part 2 has been developed to establish a mathematical calculation approach and to analyze more complex problems.

The calculation procedure is explained in **FIG. 4.**

**Example 4.** Blowdown of a compressor package can be designed to minimize the blowdown time. A choked-flow blowdown is considered (without using any restriction orifice), as indicated in **FIG. 5.** The acoustic-induced vibration issue should be checked for the following conditions.

Assumption and data:

- Settled out pressure (XV-1 and XV-2 closed), 10,000 kPag
- Both blowdown valves (DBV-1 and BDV-2) opened at the same time (no delay)
- Settled out temperature, 30°C
- Downstream temperature, –10°C
- Equivalent length of Sections A–C, 10 m
- Equivalent length of Sections B–C, 10 m
- Equivalent length of Sections C–D, 5 m
- Pipe diameter (all sections), 100 NB, Sch 120
- Pipe outer diameter (OD), 114.3 mm
- Pipe thickness, 11.31 mm
- Gas molecular weight, 16.8
- Compressibility (discharge), 0.98.

**Solution.** The main problem of this analysis is the calculation of maximum blowdown flow. Since the blowdown is from a source pressure of 10,000 kPag to atmospheric without any restriction orifice, the flow will be choked. The flowrate at choked condition must be estimated.

**Flowrate calculation.**[5,6] The volumetric flowrate for compressible fluid is calculated using Eq. 42:

$$Q = 53.64 \times Y \times d^2 \sqrt{\frac{\Delta P \times P_1}{K \times T \times S_g}} \tag{42}$$

where,

$Q$ = volumetric flow, $Sm^3$/sec
$Y$ = expansion factor
$d$ = pipe inside diameter = 0.09168 m
$\Delta P$ = differential pressure, kPa
$P_1$ = initial pressure = 10,101.3 kPa
$T$ = temperature = 303.15 K
$S_g$ = 0.58 (MW = 16.8)
$K$ = flow resistance factor
Estimation of flow resistance factor:
Total equivalent pipe length = 25 m
Friction factor (fully turbulent flow) = 0.0175
$K$ (for pipe) = 0.0175*25/0.09168 = 4.8
$K$ (entrance) = 1
$K$ (exit) = 0.5
Total $K$ = 6.3

For a compressible gas with a ratio of specific heats in the range of 1.3, the pressure ratio is estimated using Eq. 43:

$$\frac{\Delta P}{P_1} = \left\{0.9953 + \frac{0.9054}{K^{0.5}} + \frac{0.1173}{K} - \frac{0.0195}{K^{1.5}}\right\}^{-1} = 0.728 \tag{43}$$

Therefore,

$\Delta P = 0.728 \times 10{,}101.3 = 7{,}353.7$ kPa
Pressure at point D = 10,101.3 – 7,353.7 = 2,747.6 kPa
The expansion factor is calculated as (Eq. 44):

$$Y = 0.0415 \ln(K) + 0.6097 = 0.686 \tag{44}$$

The volumetric flow is calculated as (Eq. 45):

$$Q = 53.64 \times 0.686 \times 0.09168^2 \times \sqrt{\frac{7{,}353.7 \times 10{,}101.3}{6.3 \times 303.15 \times 0.58}} = 80.09\ Sm^3/s = 204{,}710\ kg/h \tag{45}$$

The Mach number is calculated as (Eq. 46):

$$Mach = 3.23 \times 10^{-5} \left[ \frac{204{,}170}{2{,}646.3 \times 0.09168^2} \right] \left( \frac{0.98 \times 263.15}{16.8} \right)^{0.5} = 1.1 \qquad (46)$$

Since the Mach number is more than 1, value of $K$ in Eq. 3 (Part 1) will be 6. The PWL design limit is determined by Eq. 47:

$$PWL_D = 173.6 - 0.125 \times \frac{91.68}{11.31} = 172.6 \text{ dB} \qquad (47)$$

PWL at source is determined by Eq. 48:

$$PWL_S = 10 log \left[ \left( \frac{10{,}101.3 - 2{,}747.6}{10{,}101.3} \right)^{3.6} \times \left( \frac{204{,}710}{3{,}600} \right)^2 \times \left( \frac{303.15}{16.8} \right)^{1.2} \right] + \qquad (48)$$

$$126.1 + 6 = 177.3 \text{ dB}$$

Since (Eq. 49):

$$PWL_S > PWL_D \qquad (49)$$

the design is inadequate. The design is to be modified to avoid acoustic-induced vibration failure.

**Modification 1.** Using the above consideration, the restriction orifice is added to limit the discharge velocity to Mach 0.7. Since the flow is not choked, the discharge pressure will be close to atmospheric. The estimated flowrate through 100-NB pipe= 4,700 kg/hr (Mach No = 0.7).

The flowrate is reduced to approximately 2% of the choked flowrate, which may be too low for this design. System adequacy can be checked for this flow condition.

The PWL design limit as calculated before = 172.6 dB.

The PWL at source calculated for this flow and pressure conditions (Eq. 50):

$$PWL_S = 10 log \left[ \left( \frac{10{,}101.3 - 101.3}{10{,}101.3} \right)^{3.6} \times \left( \frac{4{,}700}{3{,}600} \right)^2 \times \left( \frac{303.15}{16.8} \right)^{1.2} \right] + \qquad (50)$$

$$126.1 + 0 = 143.3 \text{ dB}$$

PWL at discontinuity (Eq. 51):

$$PWL_{dis} = 143.3 - 0.06 \times \frac{10}{0.09168} = 136.8 \text{ dB} \qquad (51)$$

PWL considering all sources (Eq. 52):

$$PWL_{alldis} = 10 \times \log[10^{13.68} + 10^{13.68}] = 139.8 \text{ dB} \qquad (52)$$

The overall value of PWL at discontinuity is less than the design values as well as less than 155 dB, hence the design is adequate.

The conclusion of Modification 1 can be summarized as:

- The blowdown rate will reduce significantly (the adequacy of this reduced blowdown rate is to be studied further)
- With non-choked flow, the design is adequate.

**Modification 2.** Since Modification 1 results in a significantly low blowdown rate, another design option is evaluated considering higher blowdown rate. Since the blowdown rate depends on the line size, a 400-NB line is considered.

Line size assumed, 400 NB

Line OD, 406.4 mm

Line thickness, 9.53 mm

The estimated flow for a velocity of 0.7 Mach = 84,000 kg/hr (significantly higher than Modification 1)

PWL design limit (Eq. 53):

$$PWL_D = 173.6 - 0.125 \times \frac{387.34}{9.53} = 168.5 \text{ dB} \qquad (53)$$

PWL at source (Eq. 54):

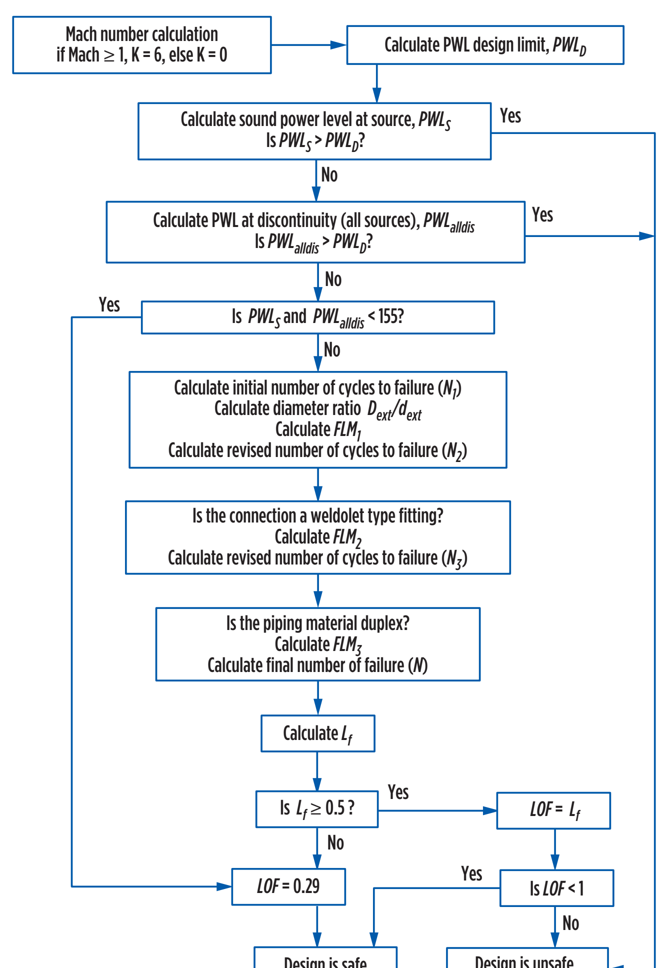


**FIG. 4.** Calculation flowchart.

$$PWL_S = 10log\left[\frac{\left(\frac{10{,}101.3-101.3}{10{,}101.3}\right)^{3.6}\times}{\left(\frac{84{,}000}{3{,}600}\right)^2\times\left(\frac{303.15}{16.8}\right)^{1.2}}\right]+126.1 = 168.4 \text{ dB} \quad (54)$$

PWL at discontinuity (Eq. 55):

$$PWL_{dis} = 168.4 - 0.06 \times \frac{10}{0.38734} = 166.9 \text{ dB} \quad (55)$$

PWL considering all sources (Eq. 56):

$$PWL_{alldis} = 10 \times \log\left[10^{16.69} + 10^{16.69}\right] = 169.8 \text{ dB} \quad (56)$$

Since the PWL considering all sources is more than the PWL design limit, the design is inadequate.

The conclusion of Modification 2 can be summarized as:

- The blowdown rate increased significantly with respect to Modification 1
- Since the PWL considering all sources is more than PWL design limit, the design is not adequate.

**Modification 3.** Modification 2 can be checked for higher line thickness as indicated in the following:

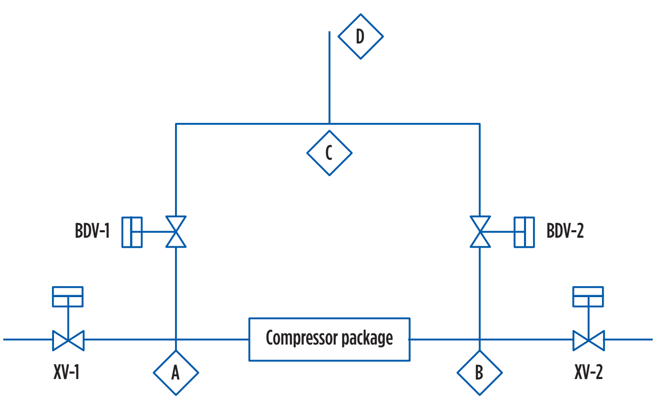


**FIG. 5.** A choked-flow blowdown is considered (without using any restriction orifice)—Example 4.

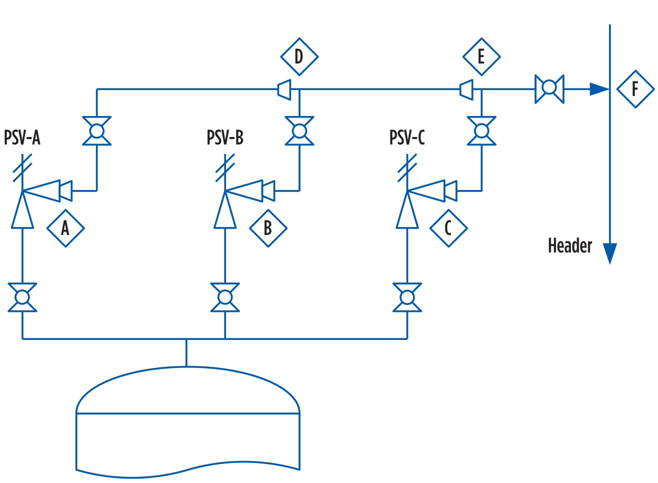


**FIG. 6.** PSV discharge configuration—Example 5.

Line size = 400 NB
Line OD = 406.4 mm
Line thickness = 16.66 mm
Line ID = 373.08 mm
Estimated flow for a velocity of 0.7 Mach = 79,000 kg/hr (little less than Modification 2)

PWL design limit (Eq. 57):

$$PWL_D = 173.6 - 0.125 \times \frac{373.08}{16.66} = 170.8 \text{ dB} \quad (57)$$

PWL at source (Eq. 58):

$$PWL_S = 10log\left[\frac{\left(\frac{10{,}101.3-101.3}{10{,}101.3}\right)^{3.6}\times}{\left(\frac{79{,}000}{3{,}600}\right)^2\times\left(\frac{303.15}{16.8}\right)^{1.2}}\right]+126.1 = 167.8 \text{ dB} \quad (58)$$

PWL at discontinuity (Eq. 59):

$$PWL_{dis} = 167.8 - 0.06 \times \frac{10}{0.37308} = 166.2 \text{ dB} \quad (59)$$

PWL considering all sources (Eq. 60):

$$PWL_{alldis} = 10 \times \log\left[10^{16.62} + 10^{16.62}\right] = 169.2 \text{ dB} \quad (60)$$

In this case, PWL considering all sources is less than the PWL design limit, but more than 155 dB. Further analysis is required to establish the design adequacy.

Calculation parameter "*a*" (Eq. 61):

$$a = 3.28 \times 10^{-7}\left(\frac{406.4}{16.66}\right)^3 - 8.503 \times 10^{-5}\left(\frac{406.4}{16.66}\right)^2 + 7.063 \times 10^{-3}\left(\frac{406.4}{16.66}\right) + 0.816 = 0.9425 \quad (61)$$

Calculation parameter "*s*" (Eq. 62):

$$s = 91.9 - \frac{406.4}{16.66} = 67.506 \quad (62)$$

Calculation parameter "*B*" (Eq. 63):

$$B = 0.9425 \times [169.2 - 0.112762 \times 67.506 - 0.001812 \times 67.506^2 + 4.307277 \times 10^{-5} \times 67.506^3] = 157.04 \quad (63)$$

Calculation of initial number of cycles to failure ($N_1$), calculate with Eq. 64:

$$\log(N_1) = 470{,}712.5155 - 63{,}075.1242 \times \log(157) + \frac{183{,}685.4368}{157^{0.5}} - \frac{575{,}094.3273}{157^{0.1}} = 9.6307 \quad (64)$$

Or (Eq. 65):

$$N_1 = 10^{9.6307} = 4.27 \times 10^9 \quad (65)$$

Correction for branch connection—Diameter ratio = 1 $FLM_1$ is calculated as Eq. 66:

$$FLM_1 = -0.07 + 0.91 \times 1 + \frac{1.32}{1} - 0.48 \times 1 + 0.065 \times 1 = 1.745 \quad (66)$$

- Correction for fittings, assuming the fitting is welding tee: $FLM_2 = 1$
- Correction of piping material, assuming the piping material is carbon steel: $FLM_3 = 1$
- Number of cycles after applying all correction factors = $7.46 \times 10^9$
- Applying a design factor of 10, the design number of cycles to failure $N = 7.46 \times 10^8$.

Calculation $L_f$ (Eq. 67):

$$L_f = -0.1303 \times \ln(7.46 \times 10^8) + 3.1 = 0.4379 \quad (67)$$

Since the value of $L_f < 0.5$, then LOF = 0.29. Therefore, the design is safe.

**Example 5.** Refer to the PSV discharge configuration in **FIG. 6** to check the possibility of AIV failure. Also, check if any modification is required to avoid AIV failure. The following design parameters (**TABLE 1**) should be used.

**Calculation of Section A to D, no additional source.** This section is designed to check the PWL at point D without considering additional sources. The source PWL will be calculated from the input parameters, therefore:

PWL at source = 0 (zero)

In this calculation, additional PWL has not been considered, therefore:

Identical discontinuity = 0
PWL additional = 0
Calculated LOF = 0.52
Health of the joint = Satisfactory

**Calculation of Sections A to D and B to D combined.** The input is i.e., the previous calculation. However, in this case additional discontinuity (Sections B to D) is added, therefore:

Identical discontinuity = 1
Calculated LOF = 0.71
Health of joint = Satisfactory

**Calculation of Section D to E (Trial 1).** The main problem with this calculation is the unavailability of input parameters to calculate PWL at source. Therefore, the value of PWL at source should be taken from previous calculations. The section flow is doubled, and a larger pipe (500 NB) has been considered for this section.

PWL design limit = 166.3 dB
PWL at source (used as input) = 166.6 dB (as calculated in previous section)
Pipe OD = 508 mm
Pipe thickness = 8.38 mm
Calculated LOF = 1
Health of joint = Failure

**Calculation of Section D to E (Trial 2).** This calculation is like the previous calculation with increased pipe thickness.

Pipe OD = 508 mm
Pipe thickness = 9.53 mm
PWL design limit = 167.2 dB
PWL at source = 166.6 dB
Calculated LOF = 0.77
Health of joint = Satisfactory

**Calculation of Sections D to E and C to E.** In this calculation, source PWL is used as input.

PWL at source (as before) = 166.6 dB
Identical discontinuity = 0

**TABLE 1.** Design parameters—Example 5

| Description | Section A to D: no additional source | Sections A to D and B to D combined | Section D to E (Trial 1) | Section D to E (Trial 2) | Sections D to E and C to E (Trial 2) | Section E to F (Trial 2) |
|---|---|---|---|---|---|---|
| Flow, kg/hr | 50,000 | 50,000 | 100,000 | 100,000 | 100,000 | 150,000 |
| Mol weight | 16.8 | 16.8 | 16.8 | 16.8 | 16.8 | 16.8 |
| Ratio of sp. heats | 1.3 | 1.3 | 1.3 | 1.3 | 1.3 | 1.3 |
| Compressibility | 0.98 | 0.98 | 0.98 | 0.98 | 0.98 | 0.98 |
| U/S pressure, kPag | 3,300 | 3,300 | 0 | 0 | 0 | 0 |
| D/S pressure, kPag | 0 | 0 | 0 | 0 | 0 | 0 |
| U/S temperature, °C | 50 | 50 | 35 | 35 | 35 | 35 |
| D/S temperature, °C | 35 | 35 | 35 | 35 | 35 | 35 |
| Section length, m | 2 | 2 | 2 | 2 | 2 | 2 |
| Pipe OD, mm | 406.4 | 406.4 | 508 | 508 | 508 | 610 |
| Pipe thickness, mm | 8.38 | 8.38 | 8.38 | 9.53 | 12.7 | 14.27 |
| Branch OD, mm | 406.4 | 406.4 | 406.4 | 406.4 | 406.4 | 406.4 |
| Branch thickness, mm | 8.38 | 8.38 | 8.38 | 8.38 | 8.38 | 8.38 |
| PWL at source, dB | 0 | 0 | 166.6 | 166.6 | 166.6 | 168.2 |
| Identical discontinuity | 0 | 1 | 0 | 0 | 0 | 0 |
| PWL additional, dB | 0 | -- | 0 | 0 | 163.8 | 0 |

PWL additional = 163.6 dB (as calculated in Section A to D)

- Trial 1
  Pipe OD = 508 mm
  Pipe thickness = 9.53 mm (as before)
  Calculated LOF = 1
  Health of joint = Failure
- Trail 2
  Pipe OD = 508 mm
  Pipe thickness = 12.7 mm (thickness increased)
  Calculated LOF = 0.71
  Health of joint = Satisfactory

### Calculation of Section E to F.

PWL at source used as input = 168.3 dB (PWL at discontinuity from previous calculation)
Line OD = 610 mm
Line thickness = 14.27 mm
Calculated LOF = 0.75
Health of joint = Satisfactory

**Takeaway.** It is clear from this work that a systematic analysis is required to establish any possibility of acoustic-induced vibration failure. The analysis must be done for each section of the system, keeping in mind the result of one section may influence the outcome of other associated sections. It is always recommended to conduct a detailed study for AIV failure where significant pressure discontinuity exists. This study is recommended not only for greenfield development but also for any existing plant. HP


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# Implement effective operational discipline programs to improve process safety performance—Part 2

Operational discipline (OD) describes human behavior in complying with required systems, every time, to consistently achieve organizational goals and overall operational excellence.[1,2] While a focus on OD is not a panacea, OD is a fundamental part of effective programs for achieving excellent performance in process safety; environmental, health and safety programs; quality; reliability; and productivity.[2,3,4]

Well-designed management systems are only the first step; the disciplined efforts of involved personnel in effectively implementing and following system requirements continuously are also needed. Failing to follow a system requirement—even just once—due to a human performance issue, inattention, complacency or other reason, can result in significant consequences, such as personal injury, environmental harm and business loss.

Part 2 here, which follows Part 1 in the February issue,[5] will discuss raising OD awareness and engagement; evaluating OD performance; identifying, prioritizing and pursuing improvement opportunities; and sustaining and renewing OD programs.

**Raising OD awareness and engagement.** When developing an OD improvement vision and specific improvement goals, it is important to share these with affected employees to raise their awareness and engagement of what OD is and why it is important, as well as to solicit broad employee input and participation on possible causes of performance issues and improvement opportunities. This can be done through (1) the introduction and discussion of OD program characteristics and goals at safety and other meetings, (2) daily face-to-face interactions and (3) other communication pathways.

OD improvement workshops (discussed in a following section) can provide focused attention to OD issues and improvement plans, helping to engage employees through their active involvement in identifying problem areas and determining priorities for future improvement activities. An important factor to emphasize, based on experience, is that OD programs promote positive discipline—such as training for an important goal or maintaining an important practice—rather than disciplinary actions, such as those related to human resource activities. With a good safety program in place and a primary goal to execute and sustain system requirements, a good message to communicate can be, "Operational discipline (OD) equals safety."

It should be understood that even well-trained employees make mistakes. Assuming a person makes safety-critical decisions every day, and assuming an error rate of 1 mistake in 100 opportunities (0.99 daily success rate), the chances of an employee completing all work tasks without error every day for a year (based on 250 workdays) would be about 8% [$(0.99)^{250} = 0.08$]. Human error must therefore be anticipated and appropriate safeguards provided to help prevent serious injuries and other significant consequences. A focus on OD is an essential—but not the only—component of effective safety programs.

Opportunities to increase the probability of success in reducing human error through improved OD and other actions, particularly for higher risk activities, are critical for improving performance in many areas. The refinery incident discussed in Part 1 (the 2005 Texas City refinery explosion and fire) provides a good example of the value of good OD; however, incidents related to the company, specific site and related facilities should be used to help personalize and engage workers, emphasizing how their work environment and safety can be affected. The recognition that good OD lowers the overall risk of safety, quality and other operational problems is basic to the understanding of the value of OD (FIG. 5).

This can be illustrated for employees in two ways to help build awareness and engagement. First, the safety triangle, shown in FIG. 6, illustrates qualitatively how significant consequences, such as serious injuries or catastrophic incidents, are often predicted by a larger number of smaller, undesirable, unsafe acts or behav-


FIG. 5. Increased OD lowers the risk of undesirable events.[6,7]

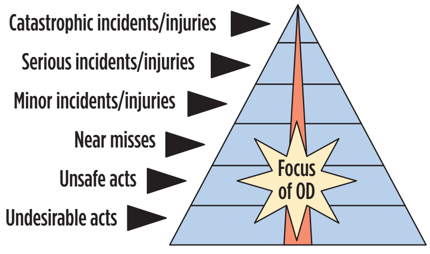


FIG. 6. The safety triangle illustrates the value of OD improvement programs.[6,7]

iors. Focusing on minimizing or eliminating problems at the bottom of the triangle helps prevent the more serious events at the top. Thus, an effective OD program works to reduce the number of unsafe acts and undesirable behaviors in the organization, as seen at the base of the safety triangle, helping to prevent more serious incidents and injuries. This is especially important for many higher-risk activities, such as working with highly toxic materials or electrical work. In these cases, the base of the safety triangle (orange triangle in FIG. 6) can be very narrow, meaning that even one unsafe act or mistake can directly lead to a serious injury or incident.

Second, the impact of OD on risk can also be illustrated in a qualitative risk matrix, shown in FIG. 7, which is often used in process hazard analyses (PHA) to estimate the risk of different hazardous scenarios. Poor OD increases risk, which typically increases the potential frequency of a hazardous event on the risk matrix since the worst-case consequences are generally used. This can increase the risk from level $R_1$ (the "perceived" risk where OD is not considered) to a higher level (the "actual" risk $R_2$), depending on the extent of the frequency increase. The resulting higher risk may be considered unacceptable in a PHA, but might not be fully evaluated if OD factors (which may impact the initiating frequency of the event and/or the reliability of safeguards) are not accounted for.

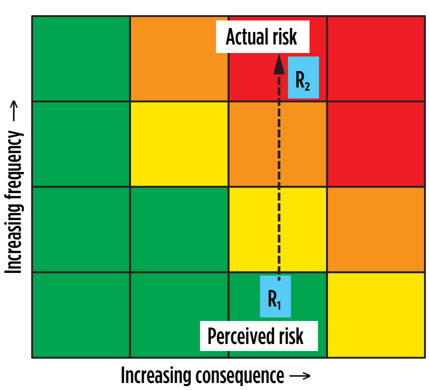


FIG. 7. Poor OD can lead to higher risk of hazardous events.[6]

Many layers of protection are typically provided to help prevent injury or significant incidents associated with process hazards, yet when an injury or incident occurs, all or many of the safeguards have failed for one reason or another. OD problems, such as a failure to follow procedures, are often some of the major reasons. In this respect, OD—or lack thereof—becomes a "common cause failure" that might impact actions taken by operators, tasks done (or deferred) by maintenance personnel, the quality of new construction work, the reliability and error tolerance in design work, etc.

Human error and related OD issues that can impact risk evaluation in a PHA or other activity must therefore be anticipated and protected against, including the use of additional risk evaluation methods that better account for failure frequencies, if needed.

## Evaluating OD performance.

Measurement is an essential activity for evaluating current performance and the possible impact and priorities of potential improvement opportunities. Sources of information for evaluating OD[6,7,8] include:

- **Incident and injury data:** Investigations of incident, injury and near-miss events focus on the identification of causal factors to learn from the event and prevent reoccurrence. Many root causes relate to OD and human performance, such as failure to follow a procedure, improper design and improper maintenance. These failures should be further investigated to identify system and cultural causes, and provide evidence of and prioritization for addressing OD issues at a site. Identifying the number and trends of incidents related to OD factors, for example, highlights the presence of and possible impacts of poor OD as well as the identification of potential improvement opportunities.
- **Audits:** Audits are conducted periodically to identify gaps in management system implementation and performance and often provide examples of poor OD. Audit findings should be reviewed to determine if they are caused by OD issues, and additional related findings can then be developed. Specific OD audit protocols[9] can be incorporated into audits to provide additional focus on potential issues.
- **Metrics:** Many safety, quality and operational metrics already being collected can provide insights on OD,[6,7,8,10] such as: incidents, injuries and near misses with OD causal factors; audit findings that relate to OD issues; overdue action items (e.g., incident or PHA recommendations); overdue scheduled activities (e.g., maintenance tests, procedure revisions); loss of primary containment events (e.g., spills); environmental permit violations; activation of safeguards (e.g., alarms or interlocks); product quality evaluations; waste levels; and equipment utilization rates. Many action items may be overdue, for example, indicating OD problems related to the timely resolution of identified improvements. In other cases, product quality problems, low process utilization and high waste levels may indicate OD causes that should be investigated further. New metrics can also be developed as needed to provide additional insights on OD performance. Trends of data

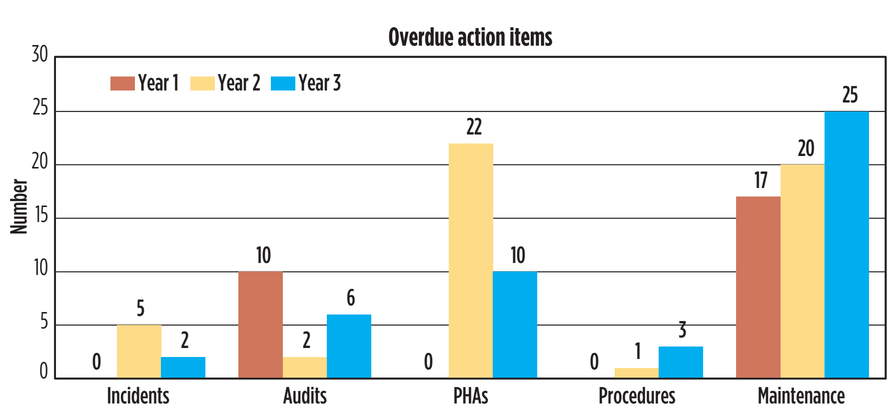


FIG. 8. OD metrics and trends can be used to identify potential issues.

over time can provide insights on whether issues are increasing or decreasing (e.g., performance is improving). An example of metrics and trend analysis is shown in FIG. 8.

- **Surveys:** OD assessments can be developed using the OD characteristics and related factors discussed earlier, using Internet survey tools or group discussions, depending on the sample size and the degree of analysis desired. Additional free-form entry of comments should be allowed when using web-based tools, and the results should be confidential to encourage honest participation. An example of a personal OD survey is shown in FIG. 9, where the X-axis could be different (1) sites, (2) groups of employees (e.g., management, operators, contractors) and (3) types of employees (e.g., work group, experience level), which can be used to examine the survey data from different perspectives. The responses in this example indicate some possible significant overall and local OD issues related to following procedures (e.g., approximately 20% of overall responses indicate a willingness to deviate from procedure, and at one site, this value is over 50%). While surveys should generally be treated as "soft" data representing employee perceptions, rather than "hard" data such as metrics related to loss of primary containment or other discrete events, survey results can be useful in helping evaluate human perspectives and beliefs that impact OD.

OD evaluation must be a continuous process to help measure progress, sustain improvements and identify potential new issues, requiring that appropriate measurement frequencies be developed. Frequent observation of work activities can also provide insights and learnings for improving OD.

**Identifying, prioritizing and pursuing improvement opportunities.** Opportunities for improving OD can be identified from employees' input and analyses of performance data, as discussed previously. A workshop approach has been used[7,11] to engage broad cross-sections of facility employees in identifying and prioritizing OD improvement opportunities. This helps raise employee awareness and understanding of OD issues, builds value for OD, engages employees in identifying problems and developing solutions, and helps the organization start prioritizing and pursuing improvements through the development of selected activities and plans. Some activities may be easy to implement and some may be much harder, requiring significant resources and time, but ultimately providing greater benefits. Generally, both types of activities should be considered, given overall resource constraints and competing activities. Prioritization can be based on factors such as overall benefit, required resources, time to complete and level of management support.

Workshops can be of varying length (typically 4 hr–8 hr), depending on site size and needs, and are a good way to engage and energize OD improvement efforts. Multiple workshops may be needed to include appropriate site personnel and, in some cases, it can be interesting and informative to view the results from workshop groups representing participants from across the site against workshop groups from different areas of a facility or different employee groups (e.g., management groups and operator groups).

While a focus on operational discipline is not a panacea, it is a fundamental part of effective programs for achieving excellent performance in process safety; environmental, health and safety programs; quality; reliability; and productivity.

An example agenda[7,11] requiring advance work to prepare relevant hard (e.g., metrics) and soft data (e.g., survey results) for review by participants can include:

- Purpose and introductions
- What is OD and why is it important?
- Breakout groups: build value for OD programs (e.g., detailed site incident review)
- Hard data review: incident, injury, audit and metrics data and trends
- Breakout groups: discuss data and identify improvement opportunities
- Soft data review: survey results
- Breakout groups: discuss data and identify improvement opportunities
- Discussion: prepare summary list of improvement opportunities and prioritize list
- Breakout groups: each group takes a different opportunity to develop a path forward
- Review group results and plan overall path forward.

The OD improvement opportunities will be specific to different locations or groups, based on the data developed and reviewed. The precursor list provided in Part 1[5] should be considered for possible activities, and some opportunities may

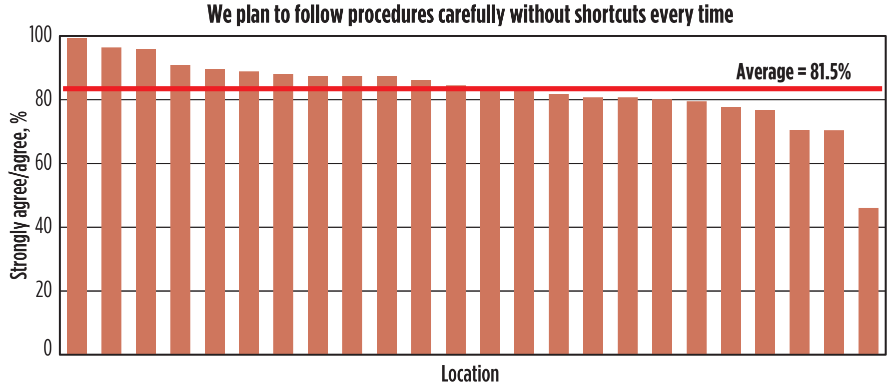


FIG. 9. OD surveys and trends can be used to identify potential issues.

overlap with safety culture and management system efforts. A list of considerations for developing possible OD improvement activities includes:

- All personnel are aware of and value OD.
- Requirements have been documented in process safety, auditing, incident investigation and other standards or guidelines (e.g., EHS, quality) to support continued focus and learning on OD.
- Procedures are current, accurate and followed consistently without shortcuts.
- Effective training is conducted for job-specific activities and troubleshooting and training is conducted on the right frequencies.
- Checklists[12] or reminder cards can be used to supplement procedures, especially for infrequent and higher-risk activities.
- Sufficient time is provided to supporting planning and conducting safe work activities.
- Factors used to evaluate the strength of a workplace[13] or safety culture[6,14,15] may influence OD more broadly.
- Some variation of "pointing and calling"[16] can help reduce mistakes, such as a required communication after completing a task.
- Complacency due to good performance is a factor and a sense of vulnerability must be reinforced.
- Sufficient and capable resources are provided to support safety-related activities.
- The organization has open and frequent communication.
- The organization promotes institutional pride through shared values.
- Behavior-based safety programs can improve commitment to safety.
- Levels of employee participation are good or show a lack of participation.
- Recognition practices are effective.
- The number of distractions present in the workplace or impacting employees.
- People are willing to ask for help or believe in stop-work guidance.
- All levels of leadership are visibly and consistently supporting safety and OD improvement efforts.
- Incident and near-miss investigations, audits and performance metrics evaluate OD causal factors, including quality and other aspects of expected operations.
- Company, site and/or area OD leaders would be useful.
- A company-wide or site-wide network to share and support OD-related activities would be useful.

Development of a documented "toolbox" or database of OD-related projects or improvement approaches—including lower priority projects that have previously been identified—can provide a useful resource for other facilities and future work.

**Sustaining and renewing OD program activities.** The pursuit of high levels of OD performance is a continuous journey that is fundamental to effective safety programs, requiring management focus and the participation of everyone at a facility. Everyone should understand why changes are being made, what the benefits are, the expected schedule, what kinds of results are expected, and what involvement and follow-through are needed.[6] Sustained progress requires leaders to embrace OD and understand that OD is not a project but a series of projects and an ongoing process. Leadership can ensure the program is sustained in the following ways:

- For each identified OD improvement project, metrics and milestones should be developed to gauge progress and help ensure that gains are sustained.
- Once one project is completed, new ones should be identified for further work and improvement.
- Existing and new requirements should be standardized based on project results and broad employee review.
- Expectations for consistent application should be clearly communicated and understood.
- Progress should be periodically reviewed by management and the results shared with employees for awareness, feedback and follow-up, as needed.
- Adjustments to projects can be made based on new information or new priorities.
- Recognition of progress should be provided, when appropriate.

Significant change is rarely easy, but implementation of a new or improved OD program can result in broad safety and operational performance gains, especially when effective management systems have already been implemented and a strong culture and supportive leadership are already in place. A review of current performance and goals can help determine if the time to get started is now. HP


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## Plasma optical emissions spectrometry analyzer

SPECTRO Analytical Instruments has introduced the newest version of its SPECTROGREEN inductively coupled plasma optical emissions spectrometry (ICP-OES) analyzer: the SPECTROGREEN TI, featuring SPECTRO's proven twin interface (TI).

SPECTRO's twin interface automatically combines both axial and radial plasma views—looking both across the plasma and from end-to-end—optimizing sensitivity, linearity and dynamic range while avoiding matrix effects like EIE. According to the company, SPECTROGREEN TI (FIG. 1) offers the highest sensitivity for trace elements, freedom from matrix interferences and good accuracy for challenging environmental matrices.

The SPECTROGREEN TI is the third and newest version of the SPECTROGREEN compact midrange ICP-OES analyzer. It delivers numerous advantages for a wide array of routine laboratory analyses: innovation and versatility; top-flight functionality; solid simplicity and ease of use; ensured durability and reliability; fast analysis for high productivity; and affordability.

Other models include the SPECTROGREEN DSOI and the SPECTROGREEN SOP.

SPECTROGREEN is ideal for routine elemental analyses in chemical/petrochemical, environmental and agronomy, consumer product safety, pharmaceutical and food applications.

SPECTROGREEN offers a competitive price/performance ratio and all models save on consumables with innovative no-purge UV-PLUS and require no added cooling, eliminating the need for expensive, breakdown-prone external chillers.

Additional features include SPECTRO's ORCA optical technology that maximizes light throughput, stability and sensitivity. SPECTROGREEN's transparency and light throughput provide high performance for UV elements, and a DSOI model, delivering twice the sensitivity of a conventional radial view across the whole spectral range. With access to the entire wavelength range via a single analysis, there is no need to choose a viewing mode or make multiple analyses—a critical advantage for high-productivity laboratories.

The analyzer's GigE readout system significantly boosts spectra processing and transport speeds for faster analysis, shorter sample-to-sample times and more samples per hour. The thermal stabilization of the optical system is achieved without sub-zero cooling. The analyzer's optimized spectral range CMOS line array technology avoids blooming while providing a fast readout and a wide dynamic range. The latest LDMOS 1,700-W generator helps analyze difficult sample matrices in lower dilutions to lower limits of detection while staying cool and trouble-free, ultimately delivering a fast warmup for higher productivity.

All sample introduction components are readily visible and accessible with its clean, compact design. Its new SPECTRO ICP Analyzer Pro operating software streamlines workflows. A modular plug-in architecture allows for individual customization. Processing speeds are ultrafast, even with large amounts of data. The analysis process is fully transparent and traceable with version and user management, plus excellent audit trail functionalities.


FIG. 1. SPECTRO Analytical Instruments' SPECTROGREEN TI.

## Range of ball valves increased

GEA has launched an addition to its range of ball valves by adding the KVplus and KVplus cryo series of top-entry ball valves designed specifically to operate with pressures of up to 100 bar. The KVplus series is capable over the full temperature range from –50°C–200°C; and the KVplus cryo series is capable over the temperature range from –196°C–100°C. This makes the KVplus and KVplus cryo series suitable for use with natural refrigerants ($NH_3$, $CO_2$), cooling brines, and non-corrosive gases and liquids, according to EN 378-1.

The new valve series completes the range—with its engineering being in high demand for many marine applications—for use in chemical factories and as a key component in heat pump technology. They are equipped with an anti-blowout stem, are fire safe tested according to ISO 10497 and API 607, and the stem-sealing system is compliant with ISO 15848 as "technically emissions free."

The design allows maintenance and inspection work to be completed without having to disconnect the valve from the line. The bi-directional seal/seat system operates irrespective of the direction of flow. When in the "closed" position, the pressure is automatically relieved via the seat system of the valve.

Operation is by manual hand lever or can be used with electric or pneumatic actuators, if required. The KVplus series valves are built within a one-piece, bolted stainless-steel housing. This provides the necessary strength to handle the most demanding applications.

## Cooling and lubricating the shaft seal on centrifugal pumps

Centrifugal pumps depend on cooling and lubrication to operate efficiently, reliably and with a long service life. The solutions commonly used for this purpose in industry are mainly constructions

made of numerous different components that are often not exactly matched to each other and require a lot of space and installation effort.

Schubert & Salzer Control Systems has now developed an out-of-the-box solution (**FIG. 2**) for this specific application. In this case, the cooling and lubrication of the shaft seals in centrifugal pumps with the operating fluid is controlled via a compact manifold. The adjustable 7010 angle-seat shutoff valve integrated into the block allows a flowrate of 10 l/hr–50 l/hr to be set.

When the pump is started, the pneumatic valve opens simultaneously, so that cooling and lubrication are ensured immediately. The liquid sensor integrated in the manifold permanently checks that coolant always flows when the pump is running. If there is no cooling, the pump is switched off to prevent damage to the shaft seal and to prevent it from getting stuck. The switch-off and switch-on functions can be carried out via the pump control or can be taken over by a PLC.

As a complete solution, the new system is not only more compact than solutions commonly used in industry, but also significantly reduces installation and maintenance costs.

The manifold is suitable for all centrifugal pumps within the specified performance range, and can be extended with manometer connections, pressure and temperature sensors, etc. Further options and intelligent, specific, modular solutions are also possible.


**FIG. 2.** Schubert & Salzer Control Systems' solution for the cooling and lubrication of the shaft seals in centrifugal pumps.

## Multi-gas detector

Concentrations of hazardous substances in the ambient air at the workplace should not exceed specified limit values. Monitoring these sometimes very low values is a demanding task. The focus of the Dräger X- act 7000, in combination with the Dräger MicroTubes for different gases and vapors, is to measure carcinogenic and toxic substances in the lower-ppb range. The range of gases to be measured is being constantly expanded. The measurement-sensitive system of the X-act 7000 is based on colorimetric chemical sensor technology and measures even the lowest ppb concentrations. It can replace conventional laboratory analysis and delivers exact reliable results directly onsite. False-positive measurement results and false alarms can be largely reduced.

The RFID tags applied to the Dräger MicroTubes contain all the calibration data that is valid for the typical period of use of one year. Complex functional tests and manual calibration procedures are no longer necessary. All possible temperature and humidity influences are already considered during factory calibration. The analyzer is explosion-proof and certified in accordance with ATEX/IECEx for zone 0 and CSA Class I, Zone 0. In addition, the system is IP54 protected against dust and splash water. It also meets the requirements of electromagnetic compatibility according to EN 61326-1.

After an automatic self-test, the X-act 7000 analysis system is immediately ready for use. The user controls the measurement task via the 3-button operation unit and the 2.4-in. color display. The measurement result, location and time can be stored in the internal data logger and read out with the Dräger CC Vision software. Power is supplied by five easily replaceable batteries. The battery capacity is sufficient for more than 10 hr of measuring and is indicated on the display. The Dräger X-act 7000 can be used with the Dräger X-am pump by connecting to it with a small connecting adapter. This makes it possible to measure carcinogenic and toxic substances in the ppb range possible even in inaccessible locations such as canals, ducts or tank facilities up to a distance of 147 ft. Since the X-am pump also has explosion protection certification for zone 0, it is ideally suited for these applications.

The X-act 7000 is manufactured by Dräger Safety AG & Co. KGaA.

## Process visualization and reporting platform

AIS Software's Version 8.2 release of the Universal Tag Locator (UTL) is an enterprise-level process visualization and reporting platform that aggregates real-time alarm, instrument, target, lab and exceedance information across a facility's stand-alone data silos into site-specific shift reports, operator dashboards and worker communications. Positive client experience has proven the value of leveraging one central communication tool for displaying operational data, reducing critical alarm floods, targeting key process values and removing the subtle barriers to process optimization.

Version 8.2 enhancement highlights include:

- Image manager
  - Insert user-captured images directly into shift reports, logs and operator instructions
  - Share images with other users
  - Link images to specific units, areas, systems or groups
- Alarm module improvements
  - Export alarm metrics data into new formats
  - Create custom line, bar and pie charts of tabular alarm data
  - Option to bulk enter external alarms
- Schedule target changes and IOW/DOW off-limit overrides
  - Schedule future target changes and overrides
- Register module expansion
  - Create blank paper forms that can be completed in Class I Div I environments
  - Improve register searches to capture more results over longer time intervals
  - Delete and archive registers in bulk.

More than 150 enhancements have been incorporated into the UTL platform. These include improvements to register field logging and column formatting, expanding status change email notifications, better support for setting targets based on PI tags, and custom handovers embedded in shift reports. HP

An expanded version of Innovations can be found online at HydrocarbonProcessing.com.